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for the refining-natural gasoline industry

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for the drilling-producing-pipe line industry

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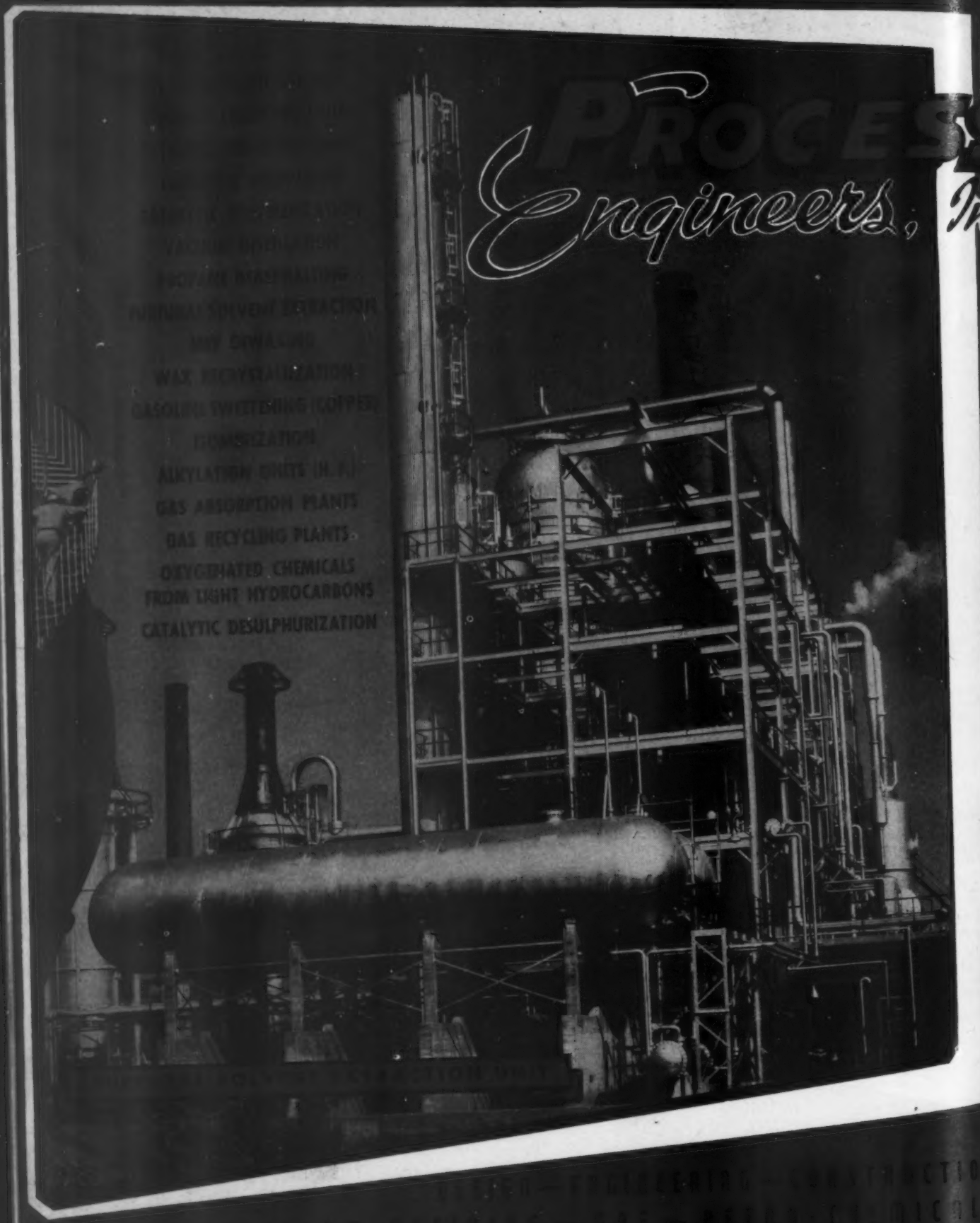
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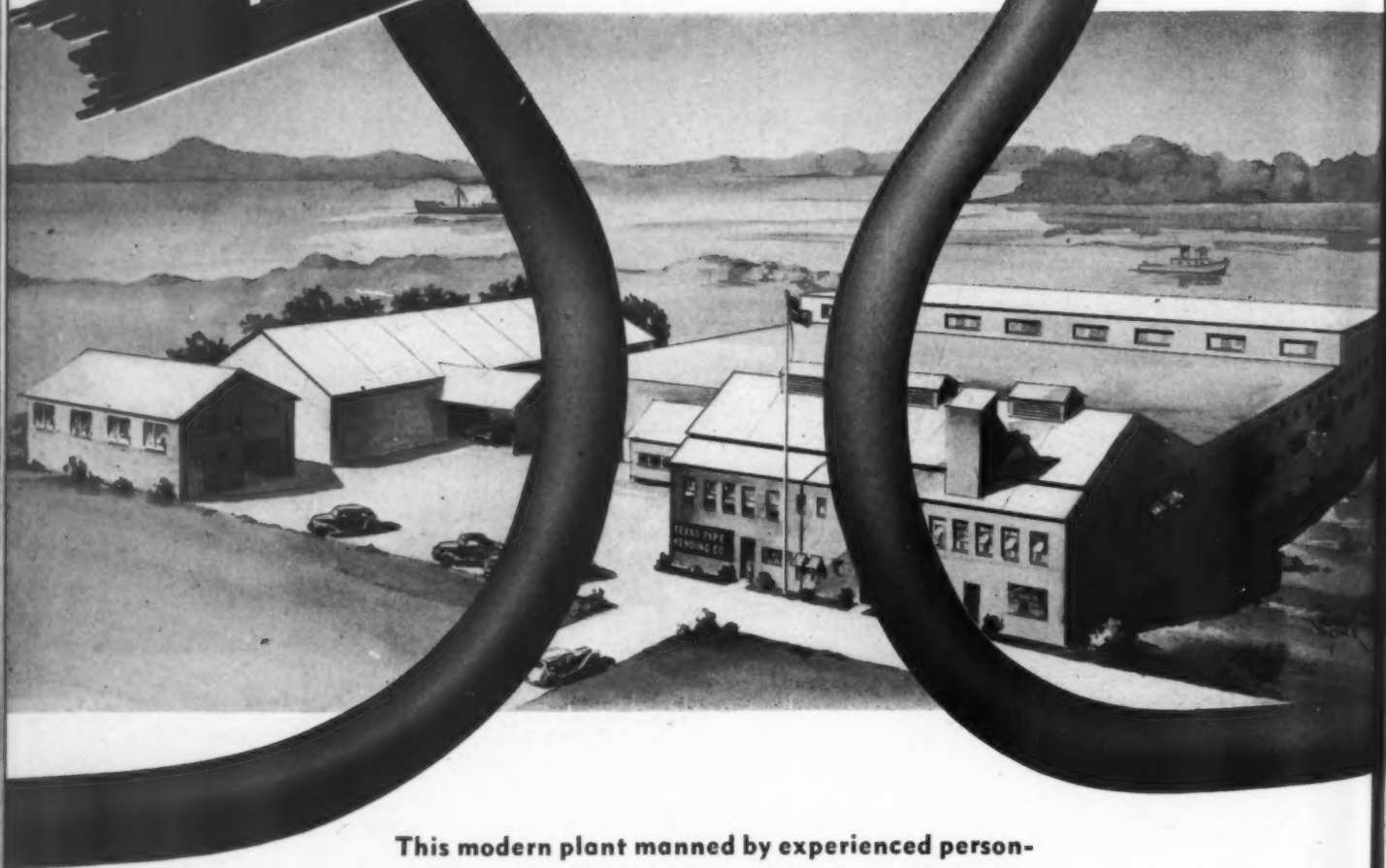

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On the Job with Arc Welding

Typical cost-cutting applications of arc welding for repair and construction of refinery equipment.

Fig. 1. Installing agitator mechanism to 70′ diameter 30,000 gallon tank for mixing ethyl in gasoline. 24″ opening in tank is flame cut in ½″ plate wall.

Fig. 2. Arc welding adapter assembly to 30,000 gallon tank shown in Fig. 1. Adapter is shop fabricated from flame cut ⅝″ plate and 24″ pipe and then held to tank with hydraulic jacks for welding with "Fleetweld 5" electrode.

Fig. 3. High pressure exchanger lining is made from 8″ wide strips of 16 gauge stainless steel weld-assembled with continuous butt welds. Assembly is then plug welded to exchanger with "Stainweld" electrode.

Fig. 4. Welding nipple to tar pot for separator. Tar pot is lined with stainless steel plate assembly plug welded to wall with "Stainweld D" electrode.

Fig. 5. Ring grooves of flanges are restored by arc welding. Two flanges are built up using "Stainweld A5-Cb" electrode with finish pass of "Stainweld D" for machinability.

Fig. 6. Welding a flange onto a 3″ mild steel pipe. Pipe is turned in a vee support for fast downhand welding.

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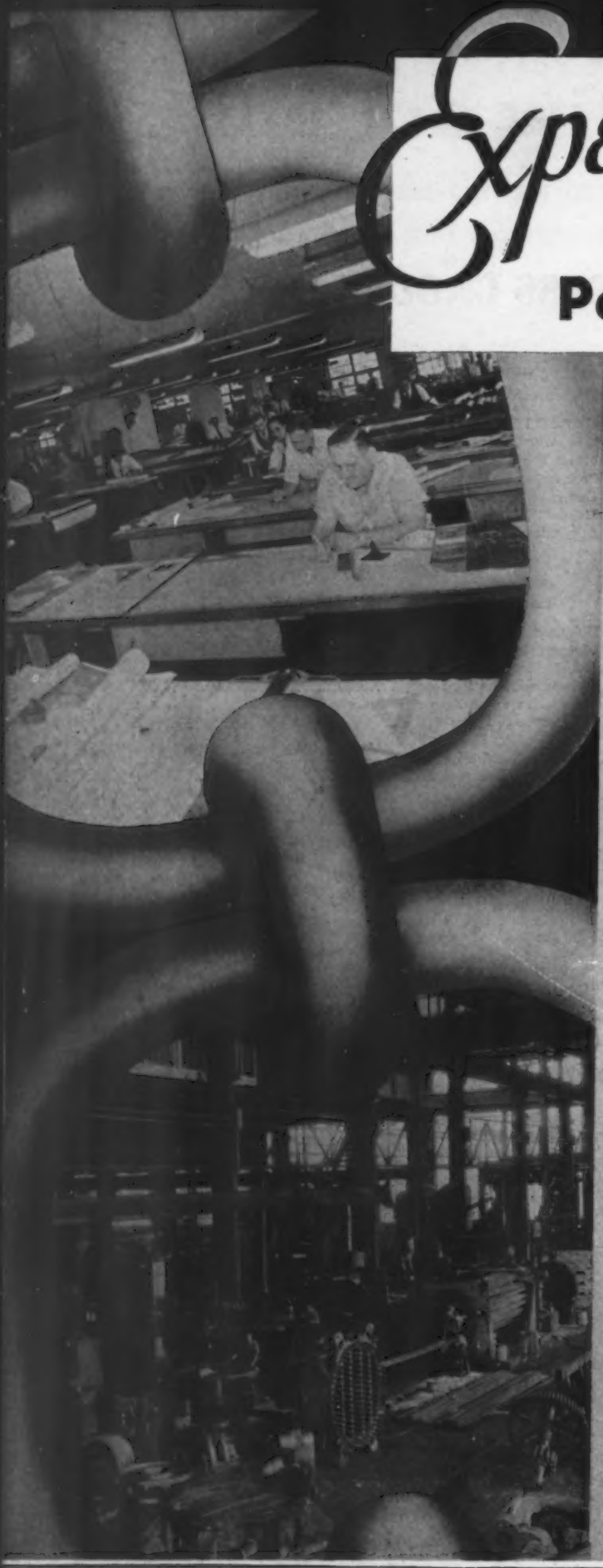

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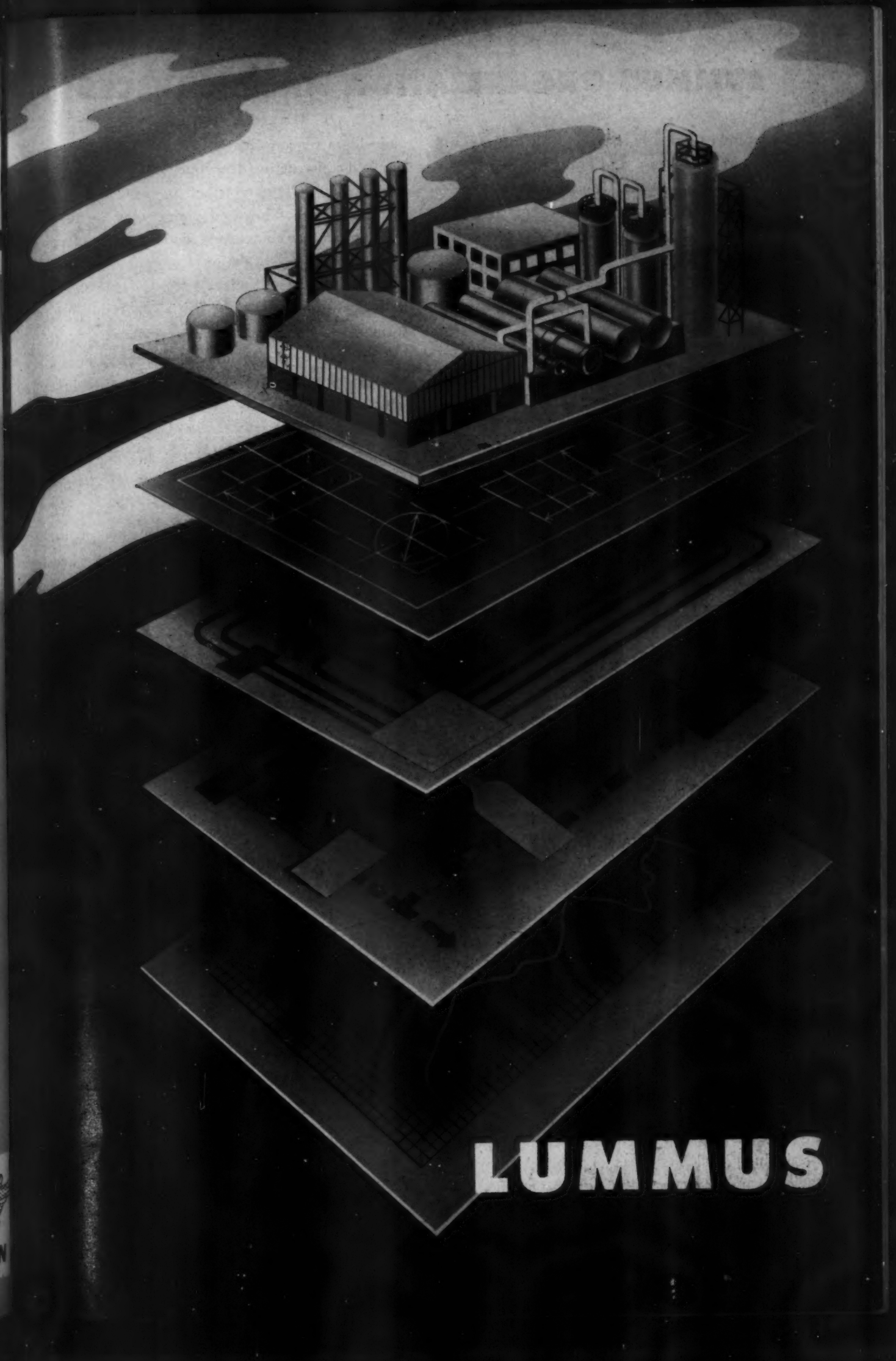

LUMMUS

LUMMUS ORGANIZATION AND SERVICES

The Lummus Company, having main offices and laboratories in New York, and branch offices in key cities of the United States, London, and Paris, is available to petroleum refiners and chemical manufacturers as a complete and integrated engineering service organization prepared to undertake anything from factual economic analysis of contemplated plant expansion to actual design, construction, and initial operation of unit process plants in these industries.

Extensive laboratory facilities are maintained for fundamental process development, charge stock evaluation, and pilot scale tests on units of varying size for such operations as distillation of all types, thermal and catalytic cracking, continuous coking, propane deasphalting, solvent refining and dewaxing, wax deoiling, and other related processes. By means of such pilot plant tests Lummus can fulfill its obligation to disinterestedly select that process or combination of processes best suited to a client's product requirements from given charging stocks. On the basis of these tests the engineering staff develops the full scale plant process and mechanical design. Material and equipment are purchased, inspected if necessary, and shipment coordinated with the requirements of the field construction department. Much of this equipment, such as all types of shell and tube heat exchange equipment and fabricated piping may be furnished by Lummus manufacturing divisions. As plant construction nears completion Lummus operating personnel conduct preliminary tests of the equipment and, finally, supervise the initial operation and performance tests for customer acceptance.

The interest of the Lummus Company in any particular installation does not end at this juncture, however, for field representatives make periodic visits to the plant keeping a record of its operating performance, maintenance requirements and long term efficiency. Should charging stocks or product requirements radically change, such check-up will enable the prompt preparation of suggested plant alterations to meet the new demands. In the last analysis, only on such long term familiarity with a plant's performance and suggestions and observations by its operating staff can future design be intelligently based.

LUMMUS BACKGROUND

Lummus has had some sixty years of experience in the design and construction of chemical plant process units and twenty-five years in the petroleum refining field. Since the birth of chemical engineering principles, Lummus has applied distillation, heat transfer, and flow of fluids in modern chemical and petroleum plant designs. Lummus has pioneered in the commercial development of a number of successful processes including lube oil solvent refining and dewaxing, selective cracking, polyforming, catalytic cracking, and wax deoiling, as well as ethylene, styrene, butadiene, phenol, and azeatrophic toluene production units, and has designed and built petroleum processing units of all types as well as many complete refineries.

In the chemical field Lummus has built over 300 units including alcohol, benzol, phenol, ethylene, butadiene, and styrene manufacturing plants. The number of plants for the manufacture of motor gasoline and 100 octane aviation gasoline from petroleum, including crude distillation, thermal cracking, catalytic cracking, polyforming and polymerization units exceed 170. In the lubriating oil field, Lummus has built over 90 process units employing solvent refining and dewaxing, vacuum distillation, propane deasphalting and clay finishing processes.

The Oil Heater Division has specialized in the intricate design of direct fired heat exchange equipment of all types and has developed a number of advantageous design features for this highly important plant element. The extensive experience of Lummus in heavy engineering and process construction, in the marshalling and supervising of large project labor forces, is reflected in its enviable records of freedom from work stoppages and rapid plant erection schedules. Lummus operations extend world-wide and include plants built in the Netherlands Indies, all through Europe and the Near East, South America, and the Orient.

GASOLINE MANUFACTURE

With gasoline production approximately 40% of crude run and quality ever on the increase, gasoline manufacturing processes represent the most important part of refinery operations.

Catalytic Cracking

Catalytic cracking was introduced in the petroleum refining field in 1938 and some one hundred plants were in operation with a total charge thruput of approximately one million barrels per day by the end of the World War II. Since then these figures have been increased in the order of 30%.

Catalytic cracking is being employed extensively for the conversion of heavy distillate fractions from crude oil into high octane number gasoline by once-through and recycle operations. This high octane catalytic gasoline has become a major constituent of motor fuel blends in many refineries and will become increasingly more important as engines designed for still higher compression ratios are brought into large scale production. Another important aspect of catalytic cracking is its ability to produce substantial volumes of home heating oil from feed stocks of higher boiling range while at the same time producing no residual fuel oil.

Virgin feed stocks for catalytic cracking units may be prepared by atmospheric and/or vacuum distillation, flashing in the presence of steam, or by propane deasphalting. Valuable additional quantities of gasoil feed stock may be derived from coking or viscosity breaking of heavy residual components of the crude which, if charged directly to a catalytic unit, would cause disproportionately large carbon deposition in comparison with the gasoline yield. Catalytic cracking units are capable of charging unvaporizable liquids as well as completely vaporized feed stocks. Single pass yields of 10 lb. RVP gasolines usually vary from 40-50% depending upon the charging stock, catalyst, and other process variables. Recycling yields of 10 lb. RVP gasoline may vary from 50-75% depending upon the aforementioned variables and the recycle ratio.

Large Gulf Coast Refinery, including Atmospheric Distillation, Combination Thermal Cracking, Catalytic Cracking, and Thermal Polymerization Units.

GASOLINE MANUFACTURE (Cont'd)

Small capacity catalytic cracking units designed for thruputs ranging from 2000-6000 barrels per day are now available and afford the small refiner the opportunity to market gasolines of competitive quality. Such units have considerable flexibility as regards recycle ratio and cracking severity, as illustrated by a recently designed unit now under construction which will produce 72% of 10 lb. RVP gasoline having a clear ASTM octane of approximately 79. Catalytic cracking in general is characterized by high recoveries of liquid products with the absence of residual fuel oil production, high octane number gasolines, substantial yields of distillate suitable for home heating oil, simplicity of operation, and high on-stream efficiencies.

Lummus has built twenty-eight catalytic cracking units designed to process a wide range of charging stocks for the production of high octane aviation or motor gasolines. Many of these catalytic cracking units are supplied with synthetic catalyst manufactured in a plant designed and erected by The Lummus Company.

Thermal Cracking

Although catalytic cracking has become a major gasoline producer in many refineries, thermal cracking units of all types continue to play an important role in refining operations. Thermal cracking units designed to process stocks selected according to boiling range and refractivity produce the highest yield of high octane number gasoline. Lummus thermal cracking units incorporating heaters with a wide degree of firing control and with separate heating and soaking sections can easily accommodate the catalytic cycle stock and heavy reduced crude cracking requirements that have been brought about by the installation of catalytic cracking facilities. Lummus viscosity breaking plants are well adapted to process heavy residues to high yields of gasoline and low viscosity-low pour fuel oil through the use of vacuum reduction and cut-back of the asphaltic residue with light cycle stock. Lummus has designed and built sixteen thermal cracking units, most of which were three-coil and four-coil combination units with visbreaking, gasoil cracking and reforming coils. The performance record of delivery time and on-stream initial operation were outstanding for these units.

Polyforming

While the commercial history of the polyform process antedates catalytic cracking, this process continues to have an important role in processing of light hydrocarbons, naphthas, and gasoils for the production of high quality motor gasolines.

The polyform unit is a self-contained plant designed to process, in addition to the naphtha or gasoil charge, propane and propylene produced in the process and also butanes and butylenes, depending upon refinery vapor pressure requirements and other considerations. The charge to the plant may also include outside propanes and butanes. The fractionating system consists of high pressure fractionator, a stabilizer and an absorber which afford the means to supply the liquefied light hydrocarbon component of the furnace charge without requiring any compression or refrigeration. The presence of these light hydrocarbons in the furnace charge makes possible substantially more severe temperature and pressure conditions than can otherwise be achieved with the result that substantially higher gasoline octanes can be obtained. At the same time the temperature and pressure conditions are such that substantially all propane, propylene, butanes, and butylenes may be converted to gasoline. Propanes and butanes produced in the process may of course be supplemented with extraneous condensibles. The large circulation of light hydrocarbons through the system exerts a stabilizing influence on the overall plant operation which contributes to the characteristic ease of operation obtained in these plants.

Polyform gasolines are characterized by high research octane numbers and exhibit excellent road performance at all engine speeds. Naphtha polyform units charging a wide variety of heavy naphthas with and without extraneous propanes and butanes achieve substantially higher yields of higher octane number gasoline than is possible in conventional reforming units operating in conjunction with catalytic polymerization facilities. In general, naphtha polyforming affords a flexible means for adjusting plant octane number and vapor pressure requirements by means of its ability to simultaneously process extraneous surplus butane-butylene and low octane number naphthas. The polyform process is also well suited for the processing of refractory catalytic cycle stocks

requiring more severe cracking conditions than can be obtained in conventional thermal cracking plants. 10 lb. RVP motor gasoline yields ranging from 60-80% have been obtained from virgin gasoils, while yields ranging from 50-70% may be obtained from catalytic cycle stocks with octane numbers ranging from 72-76 for the virgin stocks and somewhat higher for the catalytic cycle stocks.

Lummus has built twelve polyform units having a total circulation of 300,000 barrels per day, varying in size from 8,000 to 60,000 barrels per day of total circulation.

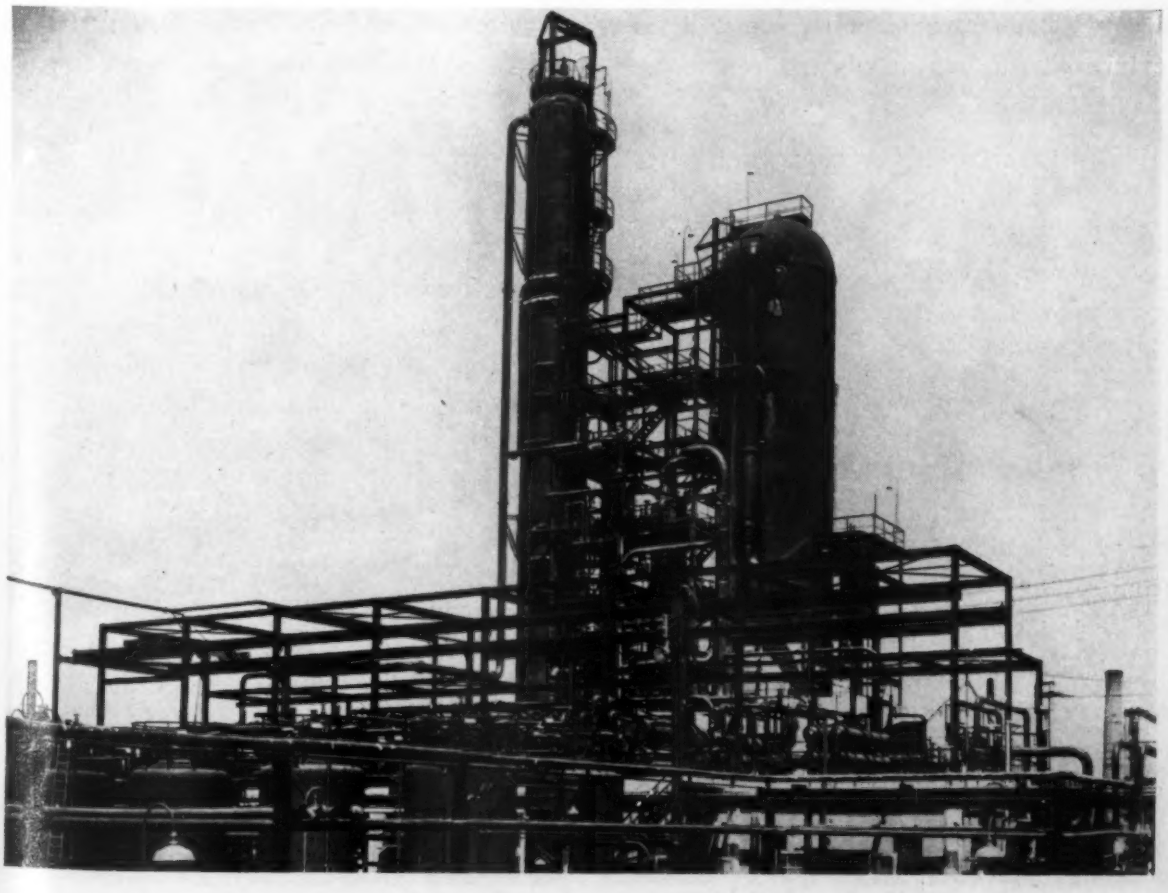
Polymerization Processes

In the field of gas polymerization, thermal and catalytic units which convert condensible light hydrocarbons to high octane polymer gaosline are available. The thermal polymerization process converts saturated as well as unsaturated hydrocarbons whereas the catalytic polymerization process polymerizes only the propylene and butylenes. Thermal polymerization gasoline yields range from 55 to 70% by weight according to the ratio of propanes and butanes, the degree of saturation and the operating conditions. Catalytic polymerization gasoline yields are substantially 100% by weight based upon the hydrocarbon disappearance, which amounts to more than 90% of the propylene and butylenes in the feed stock.

Reduced Crude Processing

The processing of reduced crude for the production of motor gasolines is largely dependent on the market demands for the various fuels. Thermal viscosity breaking, in conjunction with gasoil cracking, produces high yields of gasoline and marketable heavy fuel oil and the substitution of coking for visbreaking further reduces heavy fuel oil production. Vacuum reduction and propane deasphalting with gasoil cracking are also used for the production of high yields of gasoline with low yields of heavy fuel oil. For the preparation of charge stocks for catalytic cracking units from heavy reduced crudes, the use of visbreaking, coking, vacuum flashing, and propane deasphalting must always be weighed from the point of view of carbon deposition on catalyst and life of catalyst. The processing of reduced crudes for lubricating oils is included in the section following.

With increased use of higher speed, higher com-



Large Combination Atmospheric and Vacuum Distillation Unit.

pression automotive engines, as well as higher speed industrial machinery, the demand for better quality lubricating oils has gone hand in hand. The increasing trend towards the use of solvent deasphalting, refining and dewaxing, which had started prior to the war and which was postponed for four years, is now re-establishing itself.

A very popular lubricating oil processing system for maximum production of quality lubricating oils with minimum investment and operating costs, and maximum flexibility, includes vacuum distillation to low percent bottoms, propane deasphalting this minimum bottoms, single solvent refining the distillate cuts from the vacuum unit as well as the deasphalted oil, solvent dewaxing, and clay finishing of each of the blending stocks. Lummus has built a large number of complete lube oil plants incorporating these processes.

Vacuum Distillation

A recent Lummus complete lubricating oil plant included a two-stage vacuum unit for making raw distillate lube oil cuts which when solvent refined, dewaxed and clay finished, were suitable for blending directly into SAE grade lube oils. The features of this unit are the well fractionated cuts, low pressure drop in the towers, and the excellent color of the lube cuts. Lummus has built forty vacuum distillation units for a wide variety of services throughout the world.

Propane Deasphalting

Propane deasphalting for separating asphalt and resins from residual lubricating oil by precipitation in propane solution may be carried out in counter-current settlers or by the use of a counter-current extraction tower. Propane deasphalting makes possible a higher yield of heavy lubricating oil from a reduced crude than is possible with vacuum distillation. Aviation lubricating oils may be produced by this process in conjunction with single solvent refining and solvent dewaxing processes.

Solvent Refining

Single solvent refining for selectively removing the objectionable constituents from charge oil and thereby improving the viscosity index may be applied to a wide range of stocks from very paraffinic through mixed base to extremely naphthenic materials. Single solvent refining units show the advantage of selective refining by ability to vary the solvent dose for the different raw vacuum distillate and deasphalted cuts. Lummus has built and has under construction twenty-two lube oil solvent refining plants with a total charge oil capacity of about 55,000 barrels per day and thruputs ranging from 350 to 8,000 barrels per day.

Solvent Dewaxing

Solvent dewaxing for removing wax constituents from lubricating oil stocks has been widely accepted for the production of specification low pour oils. While this process was well established before the war, many refinements such as stepwise injection of solvent with an attendant reduction in overall solvent requirement and continuous deoiling for wax production have been added. Some advantages of the process are the narrow spread in temperature between filtering temperature and pour of pressed oil as well as between the cloud and pour points. The process operates successfully on all types of waxy oils, either raw or solvent refined, over the entire range from spindle oils through cylinder stocks, and from low melting crystalline waxes to high melting point micro-crystalline waxes. Lummus jobs include thirty solvent dewaxing plants ranging in capacity from 400 to 6,000 barrels per day and with a total oil charge in excess of 55,000 barrels per day.

Emulsion Deoiling of Wax

The emulsion deoiling process for the reduction of high oil content slack wax (40-50%) in one-pass operation to 15% and in two-pass operation to 5% is finding application in the production of crystalline waxes. This partially deoiled wax is then charged to a conventional sweater thereby increasing the wax production capacity of this plant. Installations range in capacity from 200 to 2,500 barrels per day and have all been built by Lummus.

PETROLEUM CHEMICAL PLANTS

With the realization of the possibilities of petroleum as a cheap source material for the manufacture of many chemical products, the petroleum and chemical industries have joint interests. Petroleum is now employed in the production of many chemical products such as styrene, butadiene, phenol, toluol, alcohols, acetic acid, ketones, esters, resins, plastics, and solvents. In some instances the petroleum industry has gone into the manufacture of many of these chemicals, and substantial petroleum chemical manufacturing enterprises have been developed. In other situations the chemical industry has located large chemical manufacturing establishments convenient to oil refineries with which contracts have been negotiated for the supply of petroleum raw materials.

The development of petroleum as a source material for chemicals manufacture is accompanied by the development of new large scale manufacturing technique. This results in low overall manufacturing costs and therefore lower prices to the consumer. Although to date a substantial petroleum chemical industry has already been developed, this industry is on the threshold of further expansion, and excellent opportunities await progressive refiners and chemical manufacturers.

Typical of such operations is the manufacture of ethylene for use as an intermediate in the production of chemicals and plastics. Ethylene, and by-product propylene if desired, may be economically produced in large volumes by the pyrolysis of propane, ethane, and gasoils or reduced crudes. Lummus has designed four plants for the production of ethylene from petroleum having a total output of nearly half a million pounds per day.

The Lummus Company with a broad background of experience in the design, engineering and construction of chemical plants, petroleum refineries, and petroleum chemical plants is available for the development of projects for the production of raw materials from petroleum sources and the planning of process units for the manufacture of specific chemicals.

The three 50,000 ton butadiene plants built by Lummus have already demonstrated capacities in excess of 150%. Recoveries and purity of product have exceeded design expectations. These plants are capable of making low cost butadiene equivalent to 45 to 60 million passenger car tires per year.

High Octane Aviation Gasoline Plant.

LABORATORY AND PILOT PLANT FACILITIES

The Lummus laboratories located in Bayonne, New Jersey, are well equipped with analytical, pilot plant and shop facilities and are staffed with experienced personnel for conducting all types of process investigation.

Complete analytical and small scale distillation facilities are available for crude oil, reduced crude oil and intermediate and finished product evaluations. The laboratory actively operates twenty continuous distillation systems, the largest capable of 5,000 gallons per day for atmospheric topping of crude oil, or 2,000 gallons per day for vacuum fractionation of reduced crude oils, and sixteen batch fractionation systems ranging from 25 litre capacity to a unique still capable of analyzing 1 cc.

The laboratory is equipped to establish fundamental design data for vacuum reduction, visbreaking, coking and propane deasphalting, all used for the production of cracking plant charge stocks.

Complete pilot plants are in constant use for full scale study of the manufacture of the entire range of lubricating oils. These include continuous units for atmospheric and vacuum distillation, propane deasphalting, furfural and phenol refining, solvent dewaxing and clay finishing. These pilot plants establish the yields, product quality and the design data required for the design, construction and operation of full scale commercial plants.

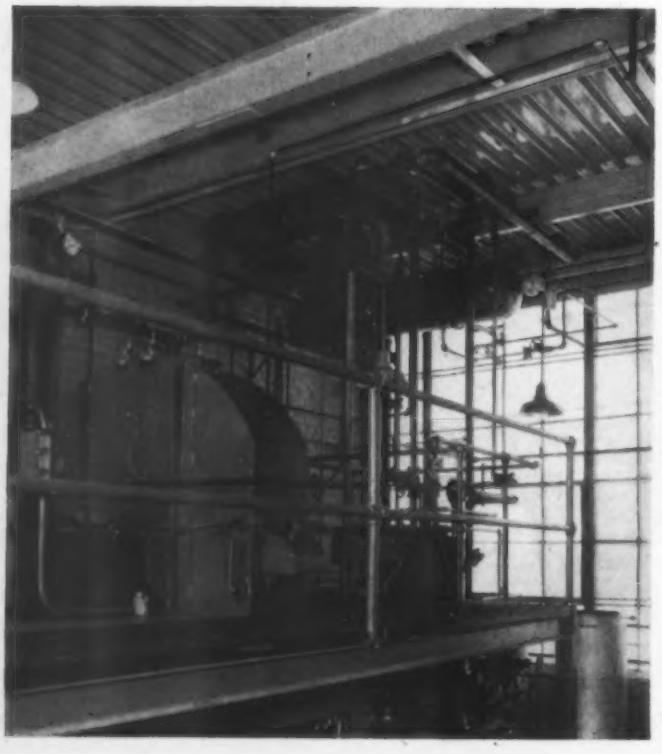
The size of these pilot plants facilitates the preparation of large samples of raw, semi-processed and finished lubes for complete evaluation of commercial lube oil characteristics and full scale engine tests.

Lummus laboratories have conducted over the past half century extensive development work for the chemical process industries for such products as alcohols, acids and esters. More recently, Lummus has projected for the petroleum chemicals industry pilot plant work for such products as styrene, butadiene, phenol, toluene, resins, plastics and solvents into full scale commercial production.

A large proportion of the laboratory equipment is fabricated in the laboratory's own shop at Bayonne. This permits an opportunity for prompt translation of the process needs into semi-plant apparatus. The translation of semi-plant to commercial scale operations is facilitated by the employment of Lummus plant operators on the pilot plant equipment. A competent mechanical crew is constantly engaged in the building of new pilot scale equipment and in its modification and maintenance.

To amplify its petroleum services, Lummus has the active cooperation of licensor laboratories.

Continuous MEK Dewaxing Pilot Plant.



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Foreword

AS THE oil industry has grown in size and complexity the single batch distillation process on which it was founded has been superceded by highly developed and specialized processes which perform the variety of separations and conversions necessary in today's refinery, natural gasoline, and petrochemical plant operations. Each improvement in product quality and each new product is usually either the direct cause or result of some development in processing methods.

"Petroleum Refiner" in this issue presents flow diagrams and descriptions for ninety processes of current importance. The processes shown serve a variety of manufacturing purposes in refining, natural gasoline, and petrochemical plants. While feed stocks, products, operating conditions, and equipment arrangements are almost endlessly variable, it may be surprising to some to note how uniform are the requirements of processes in regard to the equipment components from which they are assembled. The fact is that equipment design for the process industries has reached a high level of development and standardization, and has greatly contributed to the ease with which process designers can assemble their latest process schemes into workable commercial units.

Licensors of processes have cooperated in supplying the diagrams, process explanations, and data presented here, and the editors of "Petroleum Refiner" are deeply grateful to the personnel of these organizations for their valued assistance in this compilation. In numerous instances the licensors have prepared new diagrams and descriptions of processes which have previously been treated by "Petroleum Refiner," in order that additional data on typical operating conditions of temperature, pressure, and flow could be included. To these groups our especial thanks are extended.

In the September and October, 1947, issues of "Petroleum Refiner" there appeared brief articles concerning the use of box symbols for the recording of data on process flow diagrams. As explained in these articles, one box shape is used for temperature data, another for pressure data, others for flow quantities, and so on. It was suggested that a set of these symbols be adopted as an industry standard, because of the obvious advantages they offer in recording operating data and in showing a large amount of information on complex flow diagrams while still preserving readability.

Licensors were invited to supply flow diagrams marked thus with a typical set of operating conditions for their processes, and in many instances were pleased to comply. The response was not unanimous as some licensors considered the operating conditions of their processes too variable for adequate representation by a single set of "typical" data, or could not spare time from current work for the necessary preparation of material, or considered the major interest in their processes to be in overall results rather than details of operating conditions. A substantial percentage of the licensors, however, did elect to supply this additional data. It is the conviction of the editors that it will prove of high interest to designers, operators, and the process division of the industry in general.

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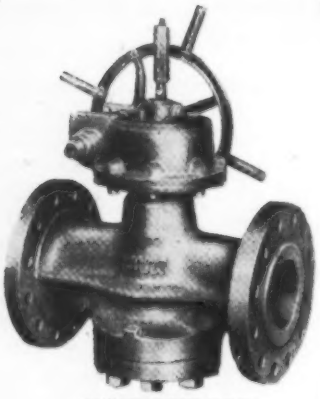

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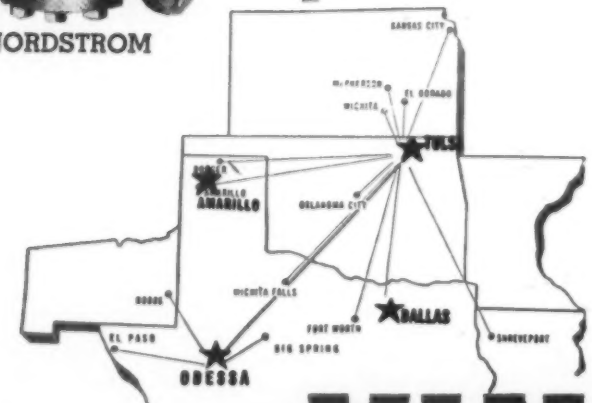

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SUPPLIES FOR THE REFINER AND NATURAL GASOLINE MANUFACTURER

Oil Distillation Processes

DISTILLATION forms the process backbone of the refining, natural gasoline, and petrochemical industry. The extensive development of catalytic conversion processes and of chemical manufacturing processes within the past decade has tended to obscure the importance of basic distillation processes and equipment. This is not a new situation. Earlier, the intense preoccupation of refiners with thermal cracking processes was such that even the cracking units often suffered in performance because of inadequate fractionation capabilities.

The prime importance of distillation is more readily apparent when it is recognized that nearly all processes require distillation for the separation of products, and that many require distillation for the preparation of feed stocks. It has been estimated that distillation facilities account for more than 60 percent of the capital investment in process plants.

Essential applications of distillation used by the process division of the oil industry include the atmospheric and vacuum distillation of crude oil for the production of straight run products, cracking unit charge stocks, and lubricants, the stabilization of naphthas, the fractionation of gases and light hydrocarbons, the separation of absorbed materials from lean oils, the separation of reaction products from recycle in conversion processes, and the separation of solvents from products in solvent treating and dewaxing processes.

Batch distillation methods used in early refineries have long since been replaced by modern continuous distillation processes utilizing bubble towers and pipestills. The batch process, involves a cycle of charge, run, and cleanout which is not adaptable to the efficient manufacture of large volumes of sharply separated products. Shell stills were installed in batteries to achieve the first continuous distillation processes. A considerable number of such units are still in use. The general adoption of the plate bubble tower and the development of the pipestill heater gave refiners a combination of equipment especially suited to the needs of the industry and culminated in the present distillation unit designs.

In the case of distillation units for light hydrocarbons, feed and products normally consist of a relatively small number of pure components and separations are measured in terms of the percentages of contaminating component left in the product streams. With heavier oils the number of components present is far too large and their identity too uncertain to permit such theoretical treatment. Instead, empirical methods of calculation are used and product separation is evaluated in terms of boiling range.

The atmospheric distillation of crude oil is by far the most important of the heavy oil distillation processes, both in respect to its basic relationship to all subsequent refinery processing and with respect to the total number of thruput of installations. Some important petroleum products require little processing other than distillation. Kerosines and diesel fuels are typical examples. At one time the bulk of gasoline manufactured was a straight run product separated from crude oil by simple distillation. Even though present day gasolines incorporate high percentages of cracked naphthas and perhaps polymer or alkylate stocks to meet anti-knock requirements, there is still a sizeable amount of straight run gasoline blended into motor fuels. Light hydrocarbons distilled from crude oils constitute an important material for the manufacture of medium grade aviation fuels. Vacuum distillation is extensively used in the rerunning of treated naphthas and lube oils, and is commercially important for use in producing straight reduced asphalts and road oils.

Crude Oil Distillation Process—Single Stage

CRUDE oil distillation is the fundamental separation process on which all subsequent refinery process steps are dependent. In numerous instances fractions of petroleum separated in simple distillation apparatus are marketed with little additional processing except possibly mild chemical finishing treatment. Fractions not directly marketed after separation still require such division as a prerequisite to further processing. Many small but quite successful refineries have a "topping" or a "skimming" unit as their principal process equipment.

The simplest and most widely used continuous crude oil distillation process employs a single bubble tower which is usually operated at or slightly above atmospheric pressure, and a pipe coil heater. More complex process units with two or more fractionating towers and heaters are used to effect savings in operating costs, to allow the vacuum distillation of reduced crudes, the removal of hydrogen sulfide along with light gasolines, and for various other sound purposes.

In a typical "topping" operation, shown on the accompanying diagram, crude oil is pumped first through heat exchange against products and reflux system streams and then through a direct fired pipe coil heater. Sufficient heat is added in the furnace to vaporize all distillate products and give a slight "overflash" at tower pressure to assure the presence of reflux liquid on the trays immediately above the flash zone. If these trays run dry sharpness of separation between bottoms and first sidestream products will suffer.

Gasoline produced overhead as a vapor is condensed for storage and to provide reflux liquid for top tower temperature control purposes. The temperature is controlled on the top tray to set the end point of the gasoline overhead product.

Other distillate products are withdrawn from the tower as liquids at appropriate tray levels. In these cases a tray is selected whose temperature will result in the desired initial boiling point for the product. The end point of a sidestream product is largely determined by the choice of initial boiling point of the next heavier product fraction, and also by the fractionating ability of the tower in question. Reduced crude is withdrawn as a residual bottoms product.

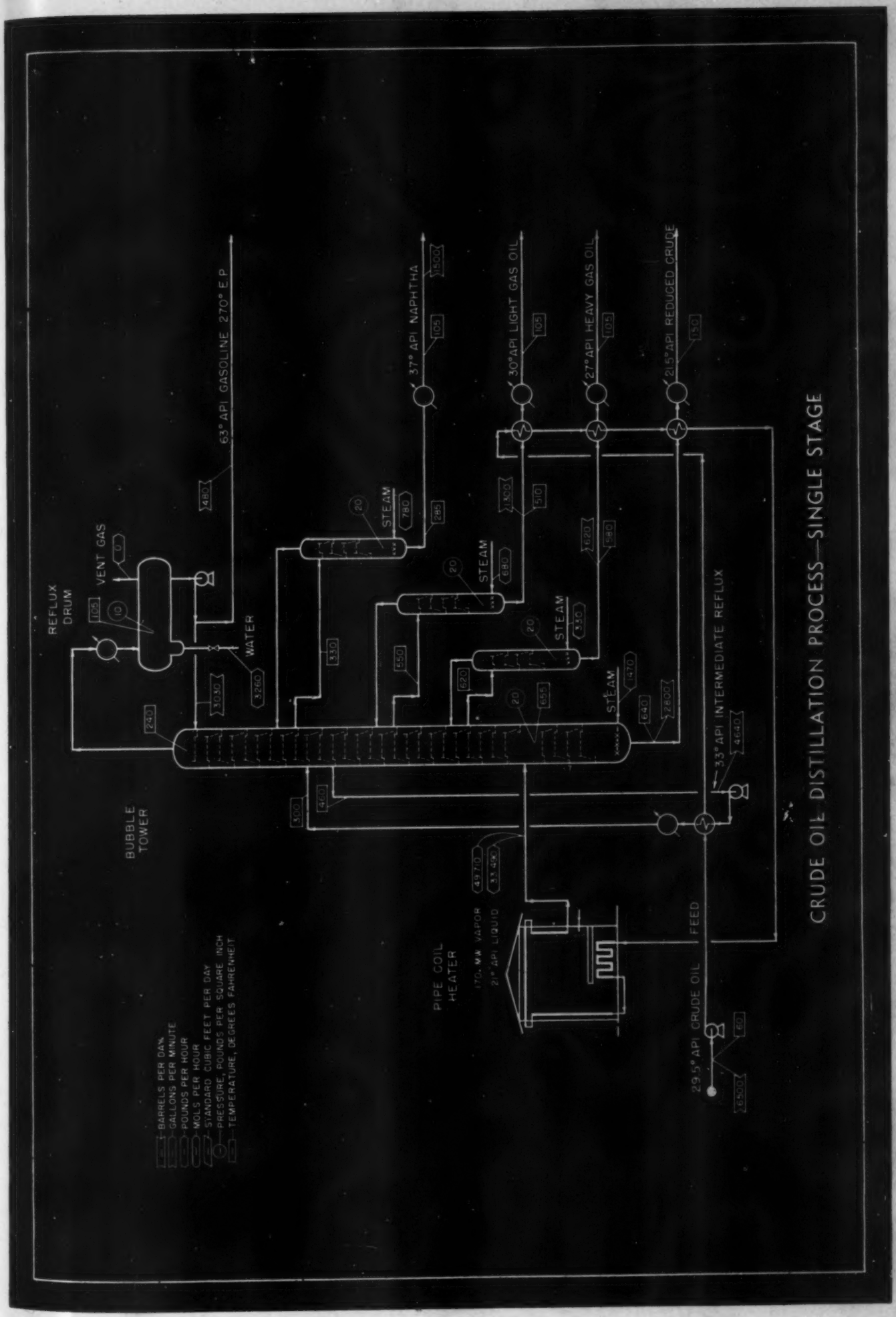
Products withdrawn as sidestreams or bottoms leave the tower in equilibrium state with lighter hydrocarbon materials which must be removed, or "stripped out" before flash and fire tests can be satisfactorily met. In general, bubble plate stripping towers are used for this purpose with steam as the stripping medium. Where especially thorough stripping is necessary, or where it is desired not to let water enter the fractionating equipment, reboilers may be substituted for steam in the strippers as a source of "boil-up" vapors.

The controlling design factor in the usual single tower topping plant is the percentage of the crude oil charge which can be vaporized for the production of distillate products.

Where maximum production of distillate is required, the degree of thermal decomposition that can be tolerated serves to limit the tower flash zone temperature. This usually sets the flash zone temperature in the range between 700° F. and 800° F. If kerosine, lubricating oils, or other sensitive stocks are processed 700° F. may be a maximum or excessive temperature, while other stocks may not be seriously damaged by temperatures that approach or even exceed 800° F. Lower total tower pressure is desirable to assist in flash zone vaporization. Since total pressure is largely composed of pressure losses between an atmospheric reflux drum and the flash zone, normally slight reduction can be made unless the tower, condenser, or piping is seriously undersized. The effect of lower tower pressure can be obtained by injecting steam to reduce the partial pressure of the hydrocarbons. Steam used for stripping tower bottoms passes through the flash zone and also assists in this manner. In general a minimum production of reduced crude is economically desirable.

Usual practice with towers of small and medium sizes is to handle the total tower reflux duty with gasoline product condensed overhead and pumped back to the tower top tray. An alternate method which sometimes saves in equipment investment utilizes a reflux condenser at dew point conditions in the overhead vapor stream. A second vapor condenser with lower outlet temperature for overhead product total condensation, and separate accumulators for both reflux and product are employed. This scheme results in high MTD's for the reflux condenser, which may have a duty several times the product condenser duty, so that condenser surface area savings may be more than sufficient to pay for the additional drum and for a pump which may be required for gasoline product movement. Another popular alternate for providing tower reflux uses two or more trays at the top of the tower for heat exchange purposes. A liquid stream is withdrawn from the lower tray of this exchanger section and pumped back through a cooler to the tower top tray. Gasoline product is condensed overhead. In large installations this scheme simplifies the overhead vapor piping and vapor condenser. Because of more favorable mean temperature difference in the reflux cooler, savings may be made in total surface requirements.

Where the overhead product is small compared to the distillate products, the ratio of internal tower reflux liquid to overhead product may be needlessly high for fractionation requirements, if all the tower reflux is introduced at the top tray. If just sufficient reflux is introduced to the tower top to supply fractionation requirements on internal reflux ratio, then an intermediate circulating reflux stream can be withdrawn, cooled, and returned to remove reflux heat from a lower section of the tower. This arrangement removes the reflux heat from the tower at a higher temperature level and so offers better opportunity for heat recovery by exchange against crude oil feed. Trays above the point of intermediate reflux introduction to a tower handle correspondingly less volume of internal reflux liquid and vapor. It is often possible to reduce the tower diameter for this upper section.


CRUDE OIL DISTILLATION PROCESS—SINGLE STAGE
29.5° API CRUDE OIL
FEED
PIPE COIL HEATER
170.- MM VAPOR
21° API LIQUID
BUBBLE TOWER
REFLUX DRUM
VENT GAS
WATER
63° API GASOLINE 270° E.P.
37° API NAPHTHA
30° API LIGHT GAS OIL
27° API HEAVY GAS OIL
21.5° API REDUCED CRUDE
33° API INTERMEDIATE REFLUX
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GALLONS PER MINUTE
POUNDS PER HOUR
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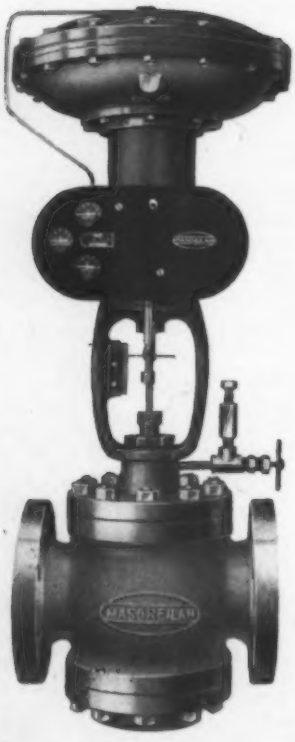

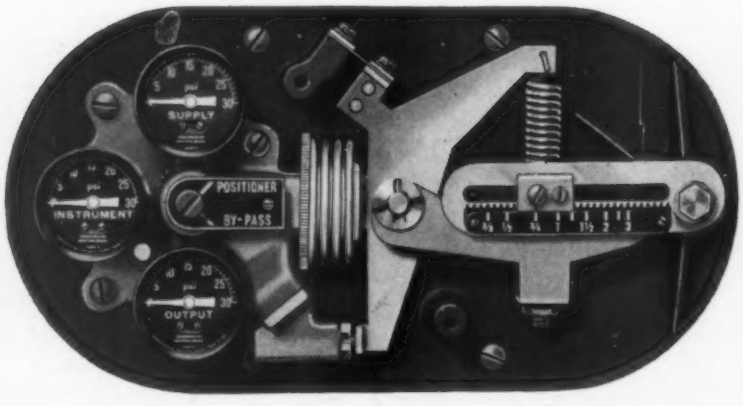

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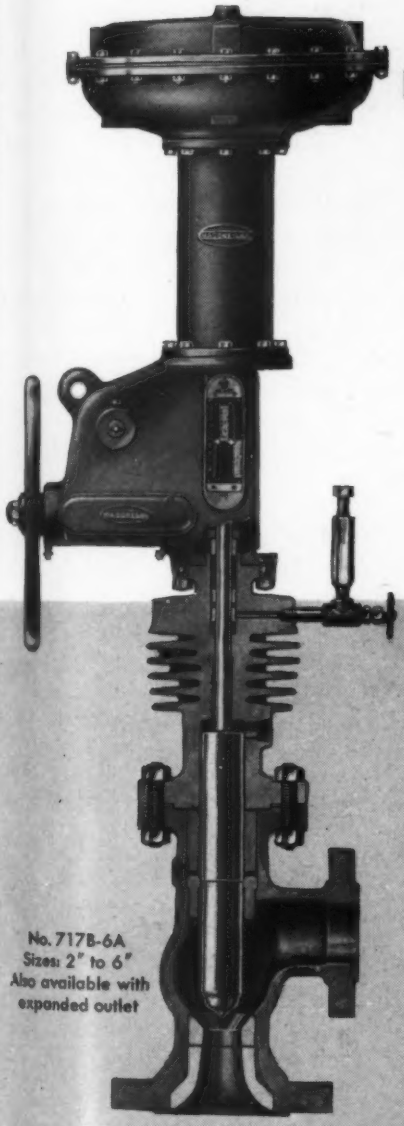

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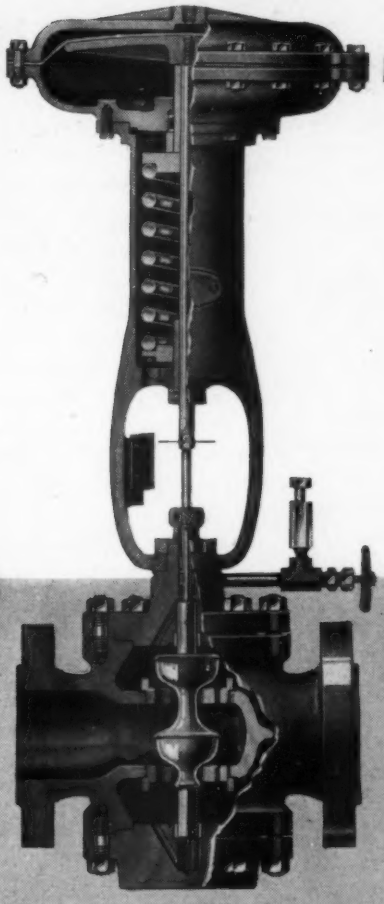

No. 37
Percentage
Parabolic
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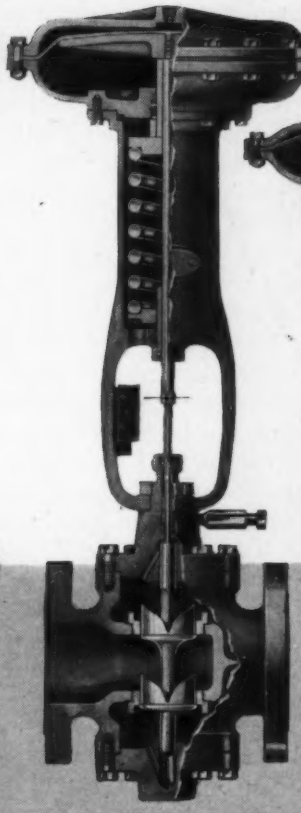
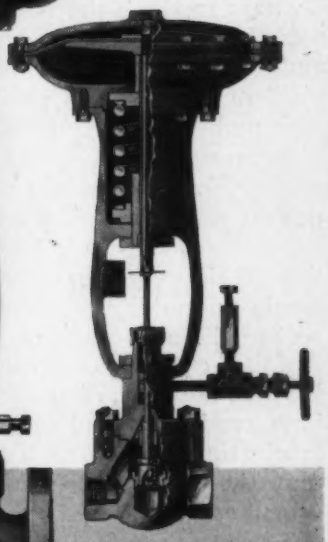


No. 38
Percentage
V-Port
Sizes: 1" to 14"



No. 37
Percentage
Piston
Orifice sizes:
⅛" to ¾"
Body sizes: ¾", 1"

No. 107
Orifice sizes:
⅛", 7/64"
Body size: ½"

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Crude Oil Distillation Process—Dual-Flo

BORN ENGINEERING COMPANY

THE refiner is often faced with the problem of processing crude oils differing widely in properties and producing products differing considerably in specifications. Some refineries have attempted to solve this problem by "batching" the crude, that is, running one type of crude for a period and then following it with another type. However, this necessitates considerable adjustment of furnace temperature and tower conditions, and results in costly delays.

The Born Dual-Flo crude distillation unit processes two crude oils separately and simultaneously. By this method, the refiner can process two crude oils of entirely different specifications simultaneously and yet take advantage of getting the maximum yield of specification product, plus operating economy and flexibility.

As an example, a refiner has two types of crude oil, one producing straight-run gasoline with a high octane rating suitable for aviation base stock and another crude with a low octane straight-run gasoline. If these crudes were blended and distilled in a single operation, it would not permit the refiner to realize the advantage of the higher octane product. Another example of the utility of the Dual-Flo unit is when the refiner has both paraffin base and asphalt base crude oils and is desirous of producing both lubricating oils and asphaltic products. By using this unit the various products could be separately obtained at the same time.

The attached flow diagram illustrates the general arrangement of the unit. The following results taken from a commercial operation of one of these units illustrates the processing:

"Charge A, a mixed crude having an average gravity of 35° API, 0.7 percent sulfur and 125-130 pounds of salt per 1000 barrels of crude is processed at the rate of 2500 barrels per day. Salt is taken out at around 200° F. and 250 psi in the settler. The overhead from this crude, a gasoline having an octane value of approximately 55, is used for blending to motor fuel. Other yields include the side cuts of tractor distillate, light and heavy gas oils and bottoms.

"A Cromwell-sand crude is used for Charge B which averages about 30.5° API gravity and contains approximately 200 pounds of salt per 1000 barrels and 0.02 percent sulfur. The charging rate for the Cromwell-sand crude is 2500 barrels per day. Gasoline from this charge stock has an octane rating of from 72 to 73 and when blended with 3 c.c. of lead produces a 90-91-octane fuel. Stoddard naphtha, kerosine and the residual product comprise the other yields from this section of the unit.

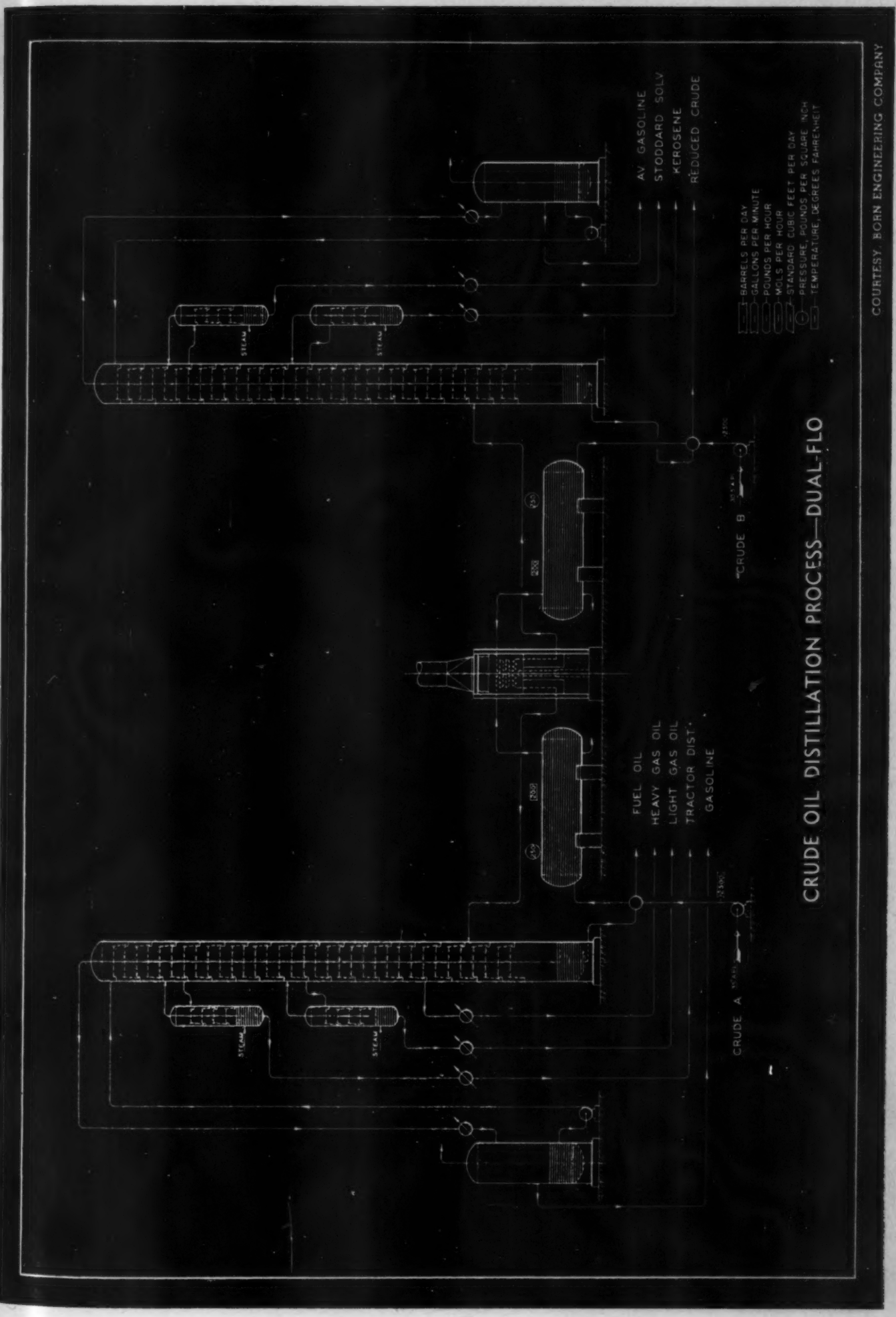
"Residual yields from both sections are mixed for a cracking stock. By combining the tractor distillate and the intermediate cuts a 350°-540° F. boiling range butadiene adsorption stock is obtained. Flexibility of the unit is illustrated in an operation when one section was rerunning pressure-distillate for an 80-octane all-purpose gasoline at the same time crude was being processed to yield an overhead naphtha."

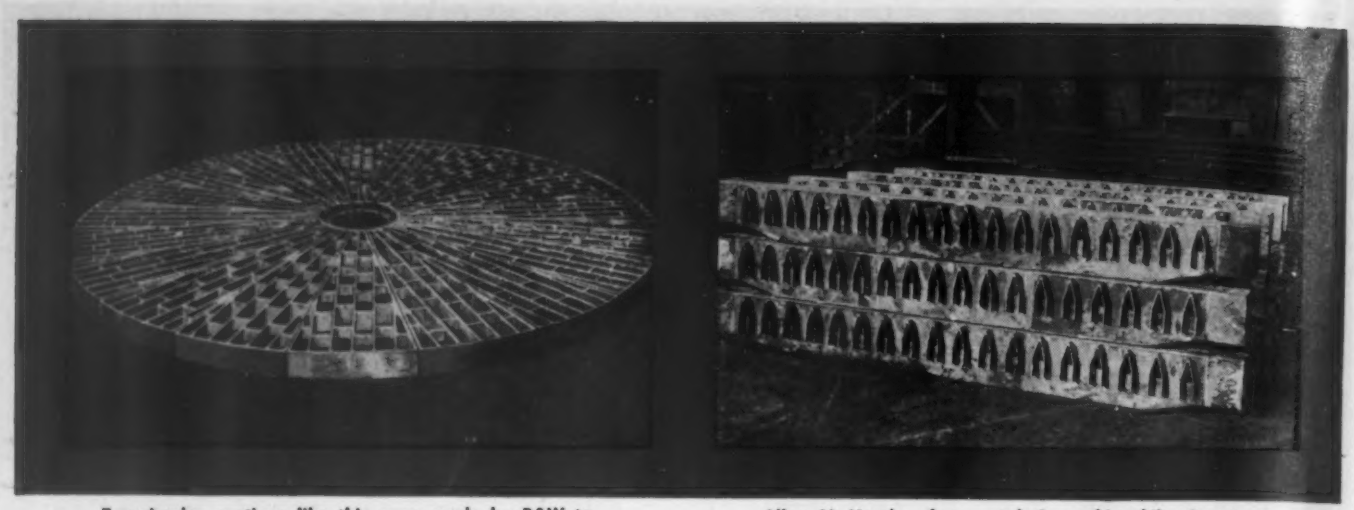
Centralization of control and other facilities common to both sections of the unit contribute to operational convenience and present a balanced plant layout. The Born Dual-Flo Heater has two combustion chambers and a single gas-duct system. The furnaces have floor-fired burners, giving close control and uniform heat distribution. Both heater coils pass from a common convection zone into their respective locations in the separate combustion chambers. Heating-coil temperatures are controlled automatically. Because each section of the heater coil is equipped with a temperature controller, the amount of heat to separate coil systems may be varied independently when desired.

Maximum flexibility of operation is accomplished by arranging the process piping so that the fractionators and the other equipment in either section of the process may be cut in or cut out as desired when running the unit on the different charges which are to be processed.

While the flow diagram shows the processing of two crude oils of different characteristics simultaneously, the same process can be applied to simultaneously heating and fractionating two products of entirely different nature, such as pressure distillate and crude oil, or distillates and naphtha. This process is especially adaptable to simultaneously heating two materials in which the yields and/or quantities are interdependent on each other.

The flow diagram on the opposite page as well as this description of the process were provided by Born Engineering Company, Tulsa, Oklahoma, designer of the unit.


CRUDE OIL DISTILLATION PROCESS—DUAL-FLO
COURTESY BORN ENGINEERING COMPANY
AV GASOLINE
STODDARD SOLV
KEROSENE
REDUCED CRUDE
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
CRUDE B
FUEL OIL
HEAVY GAS OIL
LIGHT GAS OIL
TRACTOR DIST.
GASOLINE
CRUDE A
STEAM
STEAM

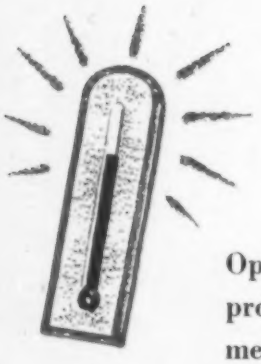


Ten circular gratings like this were made by B&W to tolerances of + ¼ inch on straightness, + ⅛ inch on length and + 1/16 inch on depth of grooves. Each had twenty 7-foot, 575-lb. sections cast from 25% Cr 12% Ni alloy.

Alloy Air Headers for a catalytic cracking kiln. Cast by B&W and assembled by welding, they have higher allowable stress at elevated temperatures, better oxidation resistance, smaller and lighter sections, longer life; cost less than fabricated headers replaced.

CAST to LAST

in high-temperature service



Vital to many different kinds of refinery equipment, B&W Alloy Castings are demonstrating outstanding ability to resist abrasion, corrosion and oxidation, and to retain great strength at elevated temperatures.

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S-83

B&W

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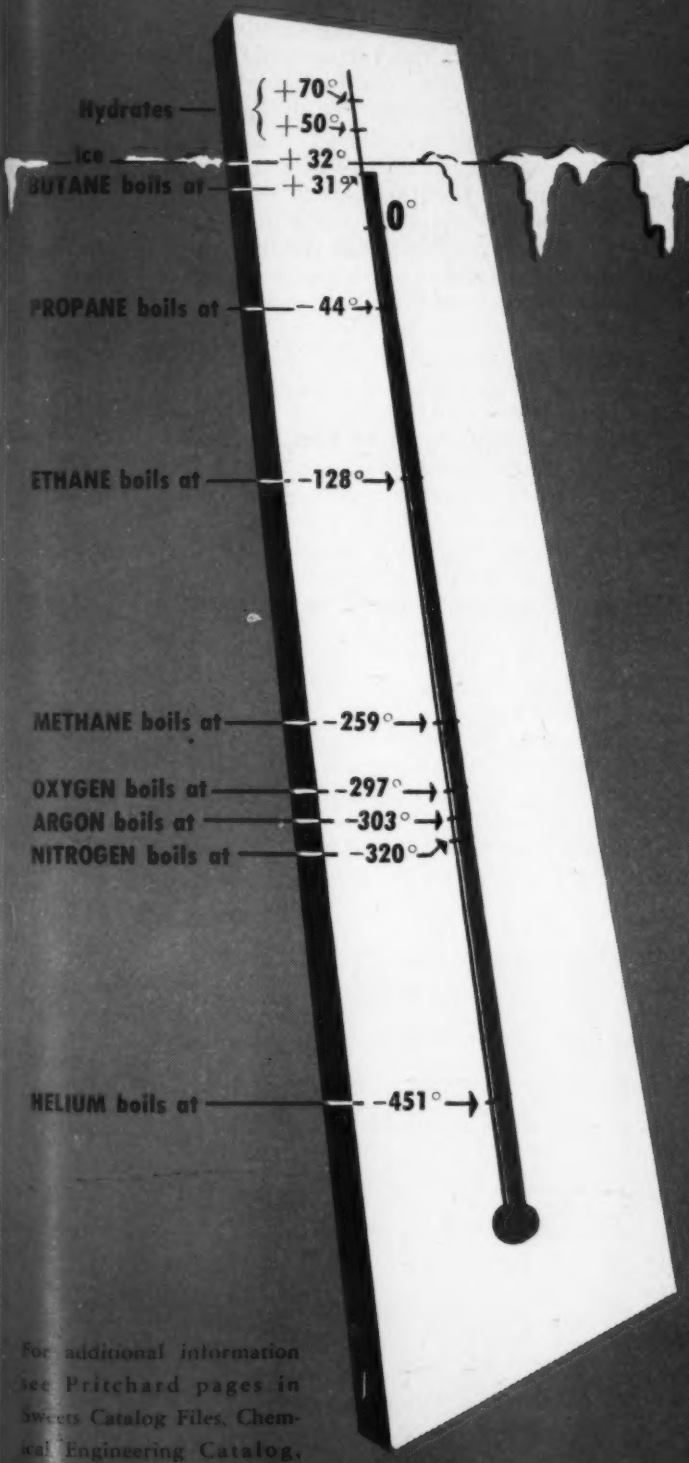
LOW TEMPERATURE



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Crude Oil Distillation Process—Two-Stage

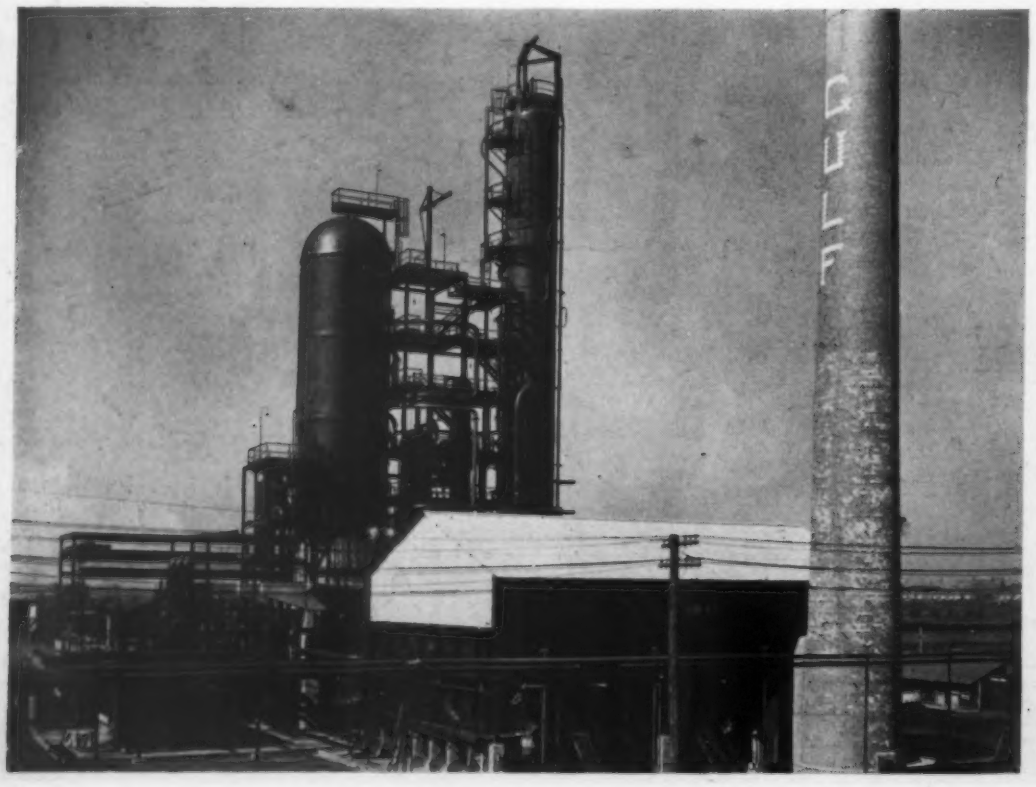
THE LUMMUS COMPANY

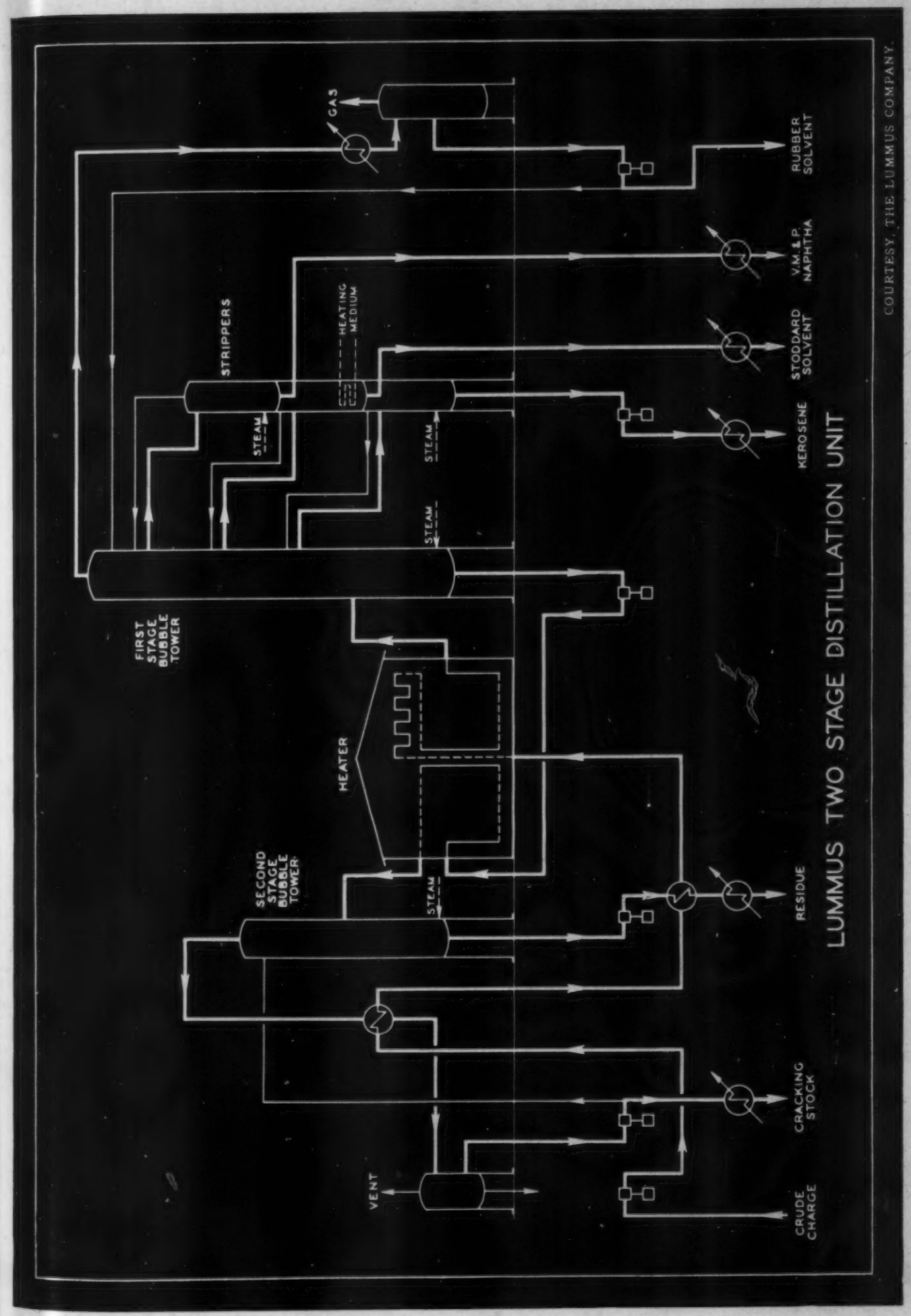
LUMMUS two-stage distillation units for the production of light gasoline, close-cut solvent naphthas, kerosine, cracking stock and fuel oil are in wide use and have excellent performance records. In these units (see flow sheet opposite) the two stages of distillation are both carried out at substantially atmospheric pressure with the temperature in the first stage being low enough to avoid the discoloration of kerosine. The temperature of the second stage is maintained at the lower limit of the cracking range to facilitate production of maximum yield of cracking stock.

The primary flashing of crude is accomplished through heat recovered by heat exchangers plus heat supplied by a coil including the convection bank and one radiant bank of a double-chamber Lummus oil heater. The products removed in the first stage include rubber solvent, varnish makers' and painters' naphtha, Stoddard solvent and kerosine. In order to have good operating control of the initial boiling point of rubber solvent, this product is condensed hot and partially weathered. The separation between rubber solvent and V.M.&P. naphtha can be easily obtained by efficient fractionation and without the use of any appreciable quantity of stripping steam. The separation between V.M.&P. naphtha and Stoddard solvent requires better fractionation than the foregoing separation, and since this latter is between the first and second side streams, reboiling of Stoddard solvent is employed. The separation between Stoddard solvent and kerosine is readily accomplished with a small amount of steam stripping applied to the kerosine stream.

For crude oils of the type for which this design is most applicable, the recoverable distillates between kerosine and fuel oil residue have no great value other than as cracking stock. For this reason, the secondary flashing operation is designed to recover a bulk distillate as an overhead product with the production of a minimum percentage of fuel oil as residue. Heat for the secondary flashing operation is supplied in the second radiant chamber of the double-chamber heater.

A feature of this plant design is the use of a partial condenser on the bulk distillate from the second flashing operation which is cooled by exchange with partially preheated crude. The temperature of the crude used as a cooling medium in this unit is closely controlled so that there is no possibility of obtaining tube-wall temperatures low enough to allow condensation of steam vapors passing through this condenser. Oil vapors are almost completely condensed and are separated from the uncondensed steam and residual oil vapors. Condensation of the resulting vapor mixture is accomplished in a spray or shower type condenser by direct contact with water. By the use of such a system the effect of corrosion resulting from mild cracking in the second heating operation is concentrated at one point, the direct-contact condenser. Because of the simplicity of design of this condenser, the cost of construction using corrosion-resistant materials is not appreciably greater than that for normal construction. Addition of neutralizing agents at this point can be done easily if desirable. Separation of water and condensate from this direct-contact condenser is carried out in a separating tank.




LUMMUS TWO STAGE DISTILLATION UNIT
COURTESY, THE LUMMUS COMPANY.
GAS
RUBBER SOLVENT
V.M.&P. NAPHTHA
STODDARD SOLVENT
KEROSENE
RESIDUE
CRACKING STOCK
CRUDE CHARGE
VENT
STRIPPERS
HEATING MEDIUM
STEAM
STEAM
STEAM
STEAM
FIRST STAGE BUBBLE TOWER
SECOND STAGE BUBBLE TOWER
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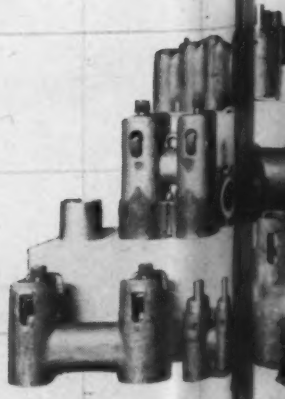

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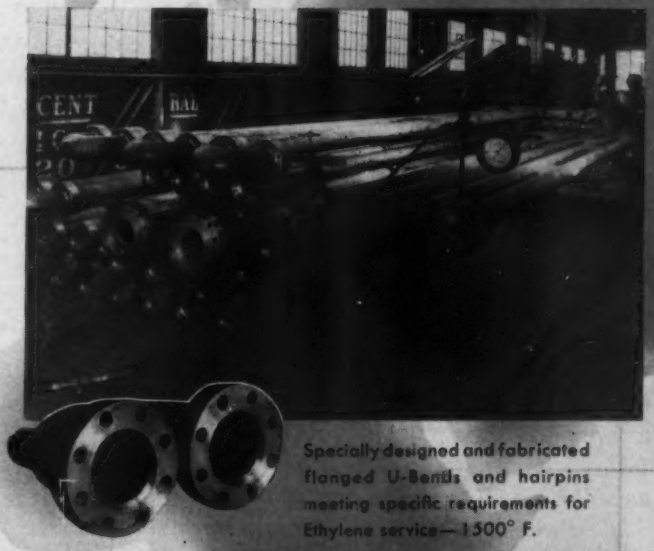

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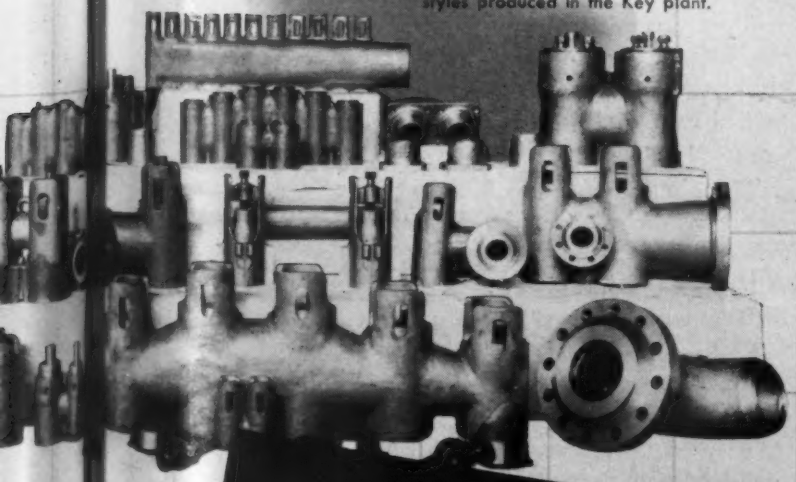

KEY Company

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Crude Oil Distillation Process—Three Stage

E. B. BADGER & SONS COMPANY

CRUDES usually are distilled under substantially atmospheric pressure for the removal of straight-run products through kerosine and light gasoil, and the reduced crude is then subjected to further distillation under vacuum for the removal of lubricating-oil fractions and the production of a heavy residuum or asphalt. These operations may on occasion be preceded by a flashing operation under mild pressure to recover light gasoline and to condition the crude for further processing.

The flow diagram on the opposite page illustrates a three-stage distillation unit designed and constructed by E. B. Badger & Sons Company, similar in general concept to the majority of modern Badger-built crude distillation units, and incorporating these three operations. Three-stage crude distillation when applied to light and medium crudes offers increased flexibility of operation; a means of combatting corrosion due to sulfur, salt and water, thereby resulting in greater on-stream efficiency; maximum yield of specification products; and improved control of fractionation.

In the first column, which usually operates at 25 to 75 psig, light gasoline is recovered overhead, and water and hydrogen sulfide are eliminated from the flashed crude. The advantages obtained in removing a light-gasoline cut in this manner before undertaking the fractionation of the intermediate cuts are as follows:

The wet crude charge is preheated in tubular exchangers at a low temperature level, minimizing corrosion difficulties from salt, water, and hydrogen sulfide. It is effectively dehydrated by flashing and stripping in the primary tower. This stripping is accomplished by reboiling the primary tower bottoms in a fired heater. Since no steam is employed for stripping, very-low-boiling products can be recovered overhead without danger of condensing water in the tower, thus minimizing corrosion from this source. Controlled amounts of light hydrocarbons can be kept in solution in the light overhead cut by operating at suitable pressures, whereas, it is impracticable to do this in the main tower at atmospheric pressure. The control of separation between the gas and light gasoline and the light naphtha can be accomplished separately and therefore more effectively. Furthermore, a properly conditioned and dried oil is obtained for charging to the fired heater, thus overcoming difficulties due to salt deposition in the heater tubes. Additionally, the top section and condensing equipment in the atmospheric tower can be designed to take advantage of the elimination of the corrosive agents in the primary tower.

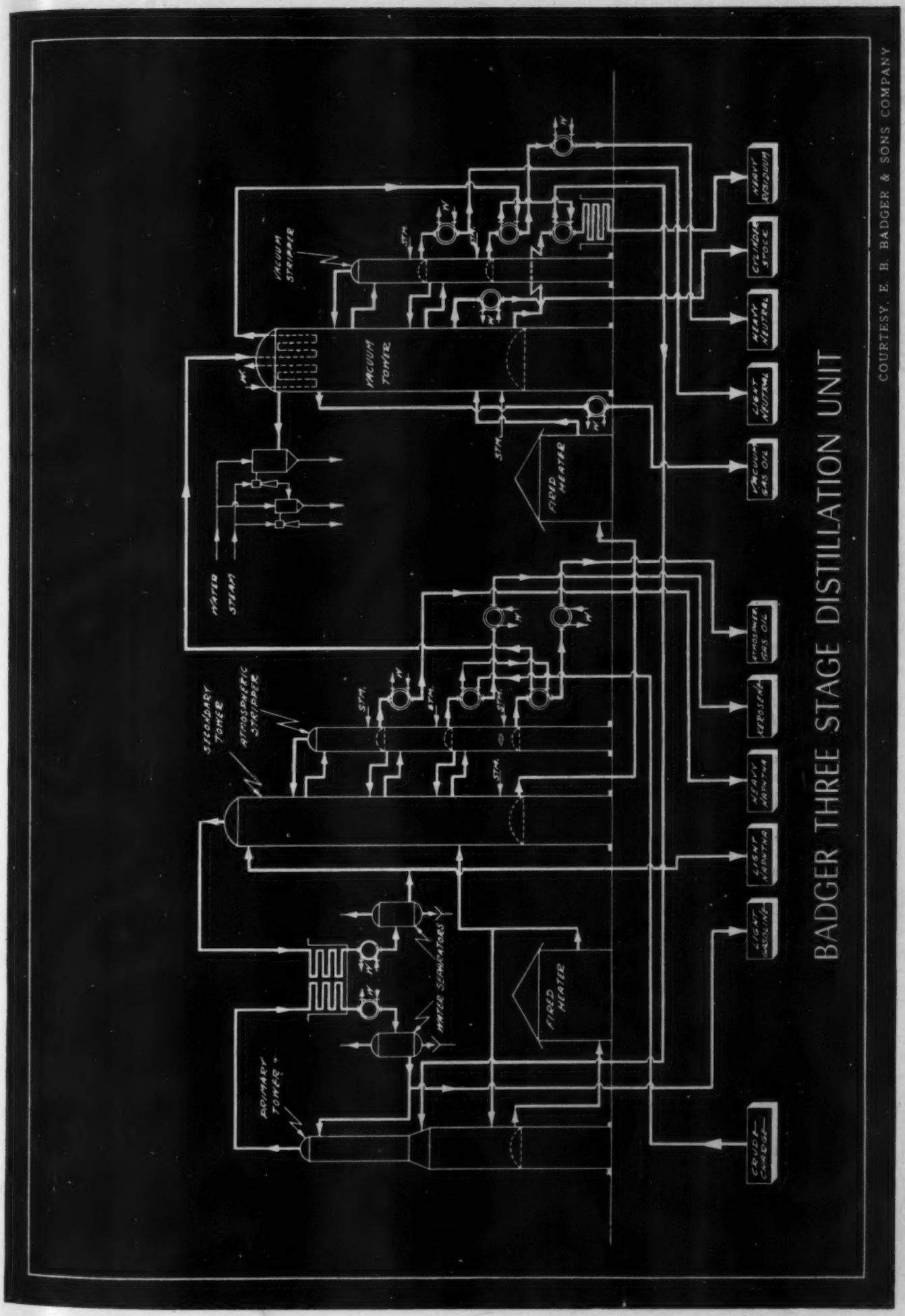
The secondary tower operates at substantially atmospheric pressure and is designed to remove light naphtha overhead, a heavy naphtha fraction and kerosine and gasoil as side streams. Steam is used for stripping the side streams and bottoms, but since the light gasoline is removed in the primary tower, the temperature level is sufficiently high to prevent condensation of water on the upper trays. The heat requirements for the atmospheric tower are provided by the same heater used for reboiling the primary tower bottoms.

In larger units or in special cases it becomes economical to provide two fired heaters, instead of one as shown in the diagram, for the first two stages. One heater circulates primary bottoms for reboiling and the other serves as a preheater for the secondary tower. Condensers can be the submerged-coil type as indicated for this unit or multitubular or direct-contact type according to conditions.

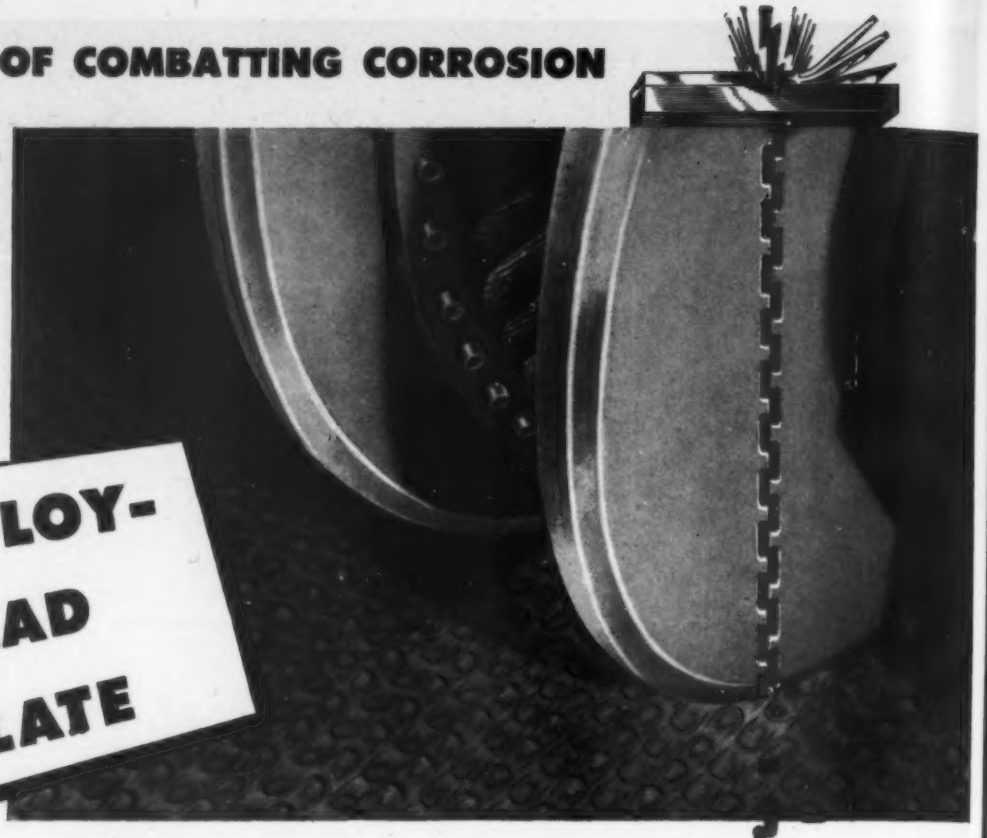
The vacuum tower produces a heavy gasoil overhead, light and heavy neutral and cylinder stock as side streams, and a heavy residuum as a bottoms product. As indicated on the flow diagram, the vacuum condensers have been mounted in the top of the tower although they may be supported above and outside the tower. In smaller units where the vapor lines are of moderate size, the condensers may be placed at a lower level for convenience. Pumps generally are used for removing vacuum distillates from the system, but in some instances where pumps are subject to severe corrosion such as by naphthenic acids in the distillate, steam lifts and barometric legs are used.

The character of the heat-exchange system is dictated by the nature of the stocks handled, and economical heat recovery varies greatly with each unit. Alternate paths of flow usually are provided for different operations.

Badger crude distillation units are designed for ease and smoothness of operation, minimum product loss while changing charge stocks, high stream efficiencies, and minimum down-town during unit turnarounds.


BADGER THREE STAGE DISTILLATION UNIT
COURTESY, E. B. BADGER & SONS COMPANY
CRUDE CHARGE
PRIMARY TOWER
WATER SEPARATORS
FIRED HEATER
LIGHT GASOLINE
LIGHT NAPHTHA
HEAVY NAPHTHA
KEROSENE
ATMOSPHERIC GAS OIL
SECONDARY TOWER
ATMOSPHERIC STRIPPER
STM.
W
WATER STEAM
FIRED HEATER
VACUUM TOWER
VACUUM STRIPPER
VACUUM GAS OIL
LIGHT NEUTRAL
HEAVY NEUTRAL
CYLINDER STOCK
HEAVY BOTTOMS

CUTS THE COST OF COMBATTING CORROSION



CROLOY-CLAD PLATE

COMBINES 9 MONEY-SAVING ADVANTAGES

The advantages listed at the right explain why B&W Croloy-Clad Plate has gained wide acceptance for fabricating process equipment requiring corrosion-, heat-, and oxidation-resistance surfaces; and protection against product contamination. It has proved to be the most economical solution to many diversified problems in refineries and chemical plants, calling for protection against destructive conditions.

Croloy-Clad Plate is produced by an exclusive B&W process which bonds a corrosion-resistant alloy cladding to a base metal. Newest machine for producing Croloy-Clad Plate is partially shown above. It represents an important advance in the art of seam-welding "clad" steels for fabrication into cost-saving process equipment.

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- Layer construction of cladding permits variation in thickness for specific requirements.
- Full corrosion resistance of lining retained by heat treatment.
- No interface carbon migration occurs between base metal and Croloy lining.
- High strength of bond permits the welding of trays, supports, and other internal fittings directly to the Croloy lining.
- Croloy-Clad Plate fabricates as easily as plain steels, and without special precautions.
- You pay no premium for the advantages of Croloy-Clad Plate.

S-84

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WORKS: ALLIANCE AND BARBERTON, O.; AUGUSTA, GA.

Thermal Conversion Processes

THE thermal conversion of petroleum fractions of low market value to other more salable products has been practiced since the early days of the industry. Operators of batch distillation apparatus discovered that prolonged heating of the heavier fractions of their charges resulted in the evolution of lighter decomposition products, and that the yield of desirable kerosene boiling range material could thus be increased. The technique was later extended to increase gasoline production.

Growing domestic demand for motor fuel was intensified by the military requirements of the first World War, and cracking processes gained in importance. The batch process was modified to semi-continuous operation. Various arrangements of tubes were added to fired shell stills to improve heat transfer characteristics, and finally the pipe still heater, developed for use with distillation units, was applied to thermal cracking operations, permitting continuous operation. A tremendous number of thermal cracking schemes were devised, and patented, and many were tried briefly in commercial installations. Only a few have survived.

Variations of the basic thermal cracking process have been developed to serve a number of useful functions. Early processes were liquid phase, but rising demand for anti-knock properties in gasoline favored the development of several vapor phase processes, somewhat more expensive in installation and operation. Thermal reforming of straight run naphthas to improve octane ratings has become common practice in the industry. The production of polymer gasoline from refinery gases through thermal conversion is practiced commercially, and various arrangements for heating gases in conjunction with naphthas and gasoils to give high yields of high quality gasoline constitute an important group of processes.

The large scale installation of catalytic cracking process plants in recent years has to a considerable degree supplanted thermal cracking as the principal conversion process for making gasoline. The catalytic units are much more expensive, however, and many of the small refiners have not adopted them even though great improvement in gasoline quality and yield can be thus obtained. In addition to this continued use of thermal cracking by the smaller refiners, one new application is the direct outgrowth of catalytic cracking operations. Considerable yields of refractory gasoil stocks are produced in catalytic cracking which are unsuitable for recycling because high coke production would result. This material can be thermally cracked to give satisfactory yields. Provision for the thermal cracking of catalytic cycle oils is, therefore, a frequent complement of catalytic cracking installations.

High temperature low pressure cracking of gases and oils to produce unsaturate and aromatic materials for subsequent chemicals manufacturing operations is widely practiced. Further development of such processes is probable but will to a large degree depend on the development of tubing materials capable of safe operation in pipe coil heaters at higher levels of temperature and pressure. Another approach to the problem of higher temperatures for pyrolysis operations is the adaptation of equipment developed for heating and handling catalysts in the continuous catalytic cracking processes. Inert materials, such as sand or pebbles, can be substituted for catalyst to serve as heat carrier medium to raise hydrocarbons to high temperatures, and to remove coke production from reaction zones. One such arrangement is included with this group of thermal cracking processes.

Shale oil distillation has received wide attention in recent years as a possible means for extending our fuel resources to meet growing demands. One method for the retorting of oil shale is included in this section because such equipment may eventually play an important role in the preparation of charge stock for cracking units.

Thermal Cracking Process

UNIVERSAL OIL PRODUCTS COMPANY

THE thermal cracking process is probably the best known of any process employed by the refining industry. It is of the widest application, both geographically and practically. Thermal cracking is in successful use in every country where oil is refined, and it is producing good gasoline from every variety of charging stock, from naphthas to the heaviest crudes and residual oils.

Universal Oil Products Company invented and developed the first cracking process that could convert heavy oils to produce maximum yields of gasoline. For more than 25 years this company's engineers have been designing, servicing and improving cracking equipment in order to make the process more serviceable to refiners of both large and small capacity.

Thermal cracking produces butylenes and propylene which can be polymerized to produce polymer gasoline of high octane number for use as a blending fluid to improve the quality of other gasolines.

The ideal equipment for thermally cracking topped or reduced crudes is centered about the two-coil cracking unit designed for optimum results when producing high-grade cracked fuel oil residuum.

The two-coil cracking unit is shown in the accompanying flow diagram.

The topped or reduced crude charge is introduced to the side of the fractionating column where it is contacted with the vapors from the flash chamber. The overhead product from the fractionating column is end-point gasoline. Furnace or tractor distillates are withdrawn as side cuts and stripped to specifications. From a lower side-cut tray a higher-boiling distillate comprising the lower-boiling fractions of both the recycle and charging stocks is withdrawn as charge to the light-oil heater. The charge to the heavy-oil heater is withdrawn from the bottom of the fractionating column, and consists of the highest-boiling fractions of both the recycle and charging stocks.

The transfer lines from the two heaters enter the top of a downflow reaction chamber. This chamber is operated without a liquid level, all products being withdrawn from the bottom through a pressure control valve.

The mixture leaving the reaction chamber enters the flash chamber in which a fuel oil residuum of the desired properties is separated.

With the development of the U.O.P. Equiflux heater and its incorporation into the two-coil selective plant described above, it is possible to obtain temperatures as high as 920° F. in the middle of the reaction chamber. With a reaction chamber operation at this temperature level the heavy charge is subjected to a far greater time-temperature action than can possibly be achieved without the use of the reaction chamber as a means of viscosity breaking. As a result, yields are improved, and plant costs are reduced because of the greatly increased cracking rates obtained at high temperatures.

Cracking residual stocks at high temperature levels has a further advantage resulting in cracked fuel oils of maximum stability and minimum sediment content. These properties are direct functions of the average temperature level of cracking, and in the modern two-coil unit the average temperature of the entire reaction chamber is greater than the maximum temperature ordinarily attained in coil-only viscosity breaking.

Similar results to a lesser degree may be obtained in a single coil and reaction chamber unit in which the charge passes through the heater with all of the recycle stock.

To illustrate the comparative results obtained in various methods of processing, the following yields are presented as representative results when charging a typical 25° API Mid-Continent reduced crude having a viscosity of 20 S.F.S. at 122° F. In considering these results, it must be remembered that not only the gravity but also the viscosity of the charge are important in determining yields.

TABLE 1

Yields from 25° API Mid-Continent Reduced Crude in Residuum Operation

	Maximum Gasoline	Maximum Furnace Oil
Percent by volume of charge:		
Gasoline, 400 E.P., 10 RVP	53.5	38
Furnace Distillate	.....	23
Fuel Oil	37.5	34
Liquid volume loss	9	5
Gas, cu. ft/bbl.	500	290
Octane number (A.S.T.M.)	69-70	68-69
Fuel Oil, vis. S.F.S. at 122 °F.	200	200
Fuel Oil, gravity, °API	7	8
Percent Catalytic Polymer Gasoline, 10 RVP	5	3.5
Percent Total Gasoline, including Polymer	58.5	41.5
Octane Number of Total Gasoline	71-72	70-71

It should be noted that the cracked gasoline yields of Table 1 are reported on the basis of 400° F. end point and 10-pound Reid vapor pressure. On the basis of total butane-butylene retention in the gasoline the yields would be increased approximately 4 percent and 3 percent, respectively. The yields of catalytic polymer gasoline shown in the table are on the basis of polymerizing 80 percent of the propylene and butylenes formed in the cracking operation.

GAS TO POLYMER-IZATION UNIT
ABSORBER
STABILIZER
STABILIZED DISTILLATE
TO STORAGE
CHARGE
FRACTIONATOR
FLASH CHAMBER
RESIDUUM FLASH COLUMN
RESIDUUM
REACTION CHAMBER
HEATERS
LIGHT OIL
HEAVY OIL
THERMAL CRACKING PROCESS
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Thermal Cracking Processes

THE LUMMUS COMPANY

NOW that more than a million barrels of per stream day capacity of catalytic-cracking units are drawing upon gasoils, which formerly were charged to thermal processes, the selective thermal cracking of heavy crude residues and refractory catalytic gasoils is becoming more and more important. While the solution in most plants involves adaptation of existing units to the new requirements, rather than the installation of new facilities, there still remain situations in which it is desirable to install new thermal-cracking facilities.

Selection of proper equipment often is difficult because of the availability of many types of units, as well as because of special conditions and the specific problems involved. The important factors which determine the performance and economic value of a projected thermal cracking plant may be summarized as:

Product yield and quality, initial investment, service factor, fuel and power economy, flexibility, maintenance and turnaround cost, time required for construction and time lapse between initial starting-up and acceptance.

A comparison on the basis of the above significant factors will indicate that Lummus units embody the best proven principles of the art aplied in the way best suited for the specific conditions of the refiner. The units are designed and erected under existing license agreements but they incorporate novel features and refinements of design which assure superior performance.

The yield efficiency and product quality to be expected from these cracking plants can be substantiated by commercial acceptance tests. Lummus combination cracking units are designed for the optimum segregation of furnace charge stocks and their independent selective cracking under the most suitable conditions. In some units stocks are segregated for selective cracking not only according to boiling range but also according to source or refractivity.

In order to crack selected charging stocks under optimum conditions it is necessary that the heaters be accurately designed to maintain optimum conditions of temperatures, shape of heating curve, cracking time and rate of heat transfer through the furnace. In the ordinary cracking units these limitations preclude flexibility as to charging stocks and product specifications. Lummus units meet the conditions of both flexibility and efficient selective cracking with carefully planned fractionation and recovery systems and by novel cracking-heater design. In Lummus furnaces the preheating and soaking sections are independently-fired chambers. This allows the operator a wide range of variation in cracking conditions without change in equipment and without sacrificing constancy of control for any given condition. This feature, combined with the unique method of handling combustion gases in the heating chamber makes Lummus cracking heaters outstanding in the field.

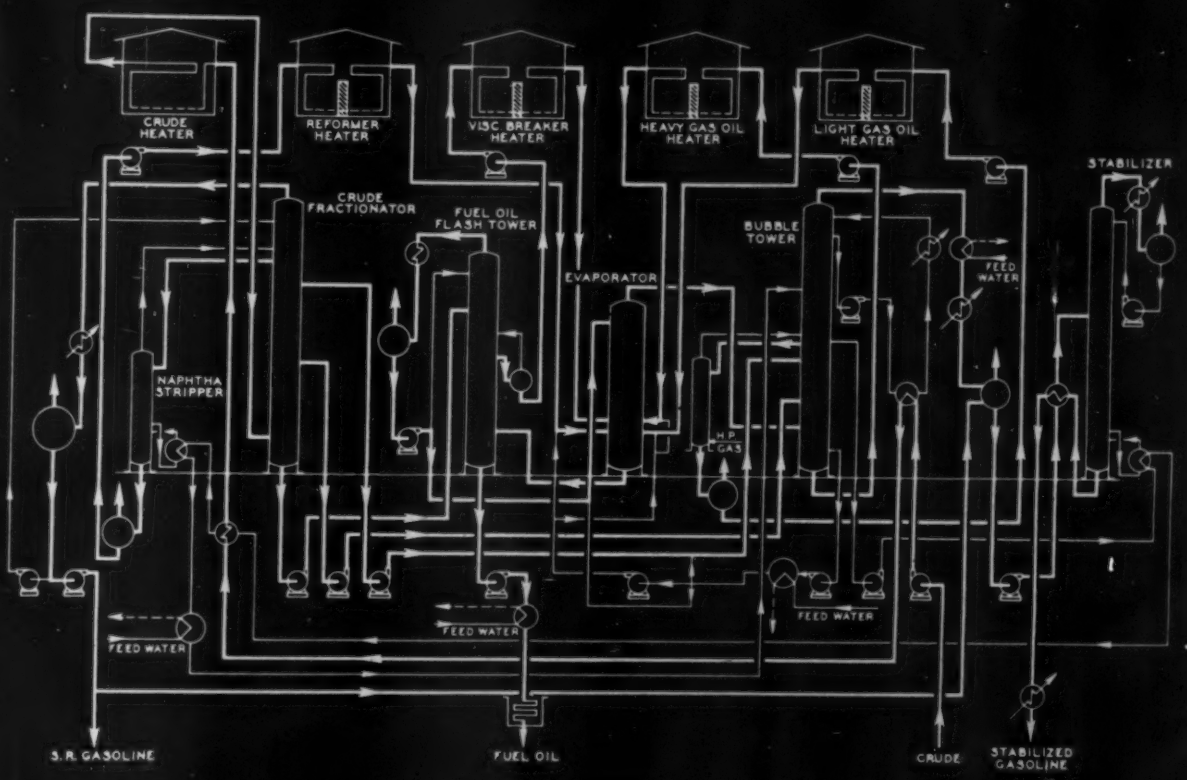
The design of the fractionating equipment of Lummus combination units reflects experience of over 40 years in the distillation field. Most combination cracking units have an excess of recoverable heat within certain temperature ranges. This heat has been utilized by generating high-pressure steam. For example, one unit generates 280,000 pounds per day of 125-pound steam. The total steam consumed by this unit is only 373,000 pounds per day. Furnace-charge pumps in this case were driven by gas engines.

Symmetry of equipment arrangement and refinements in layout pay dividends in operating convenience and rapid, inexpensive turnarounds. These features can best be appreciated by a visit to one of the units in operation.

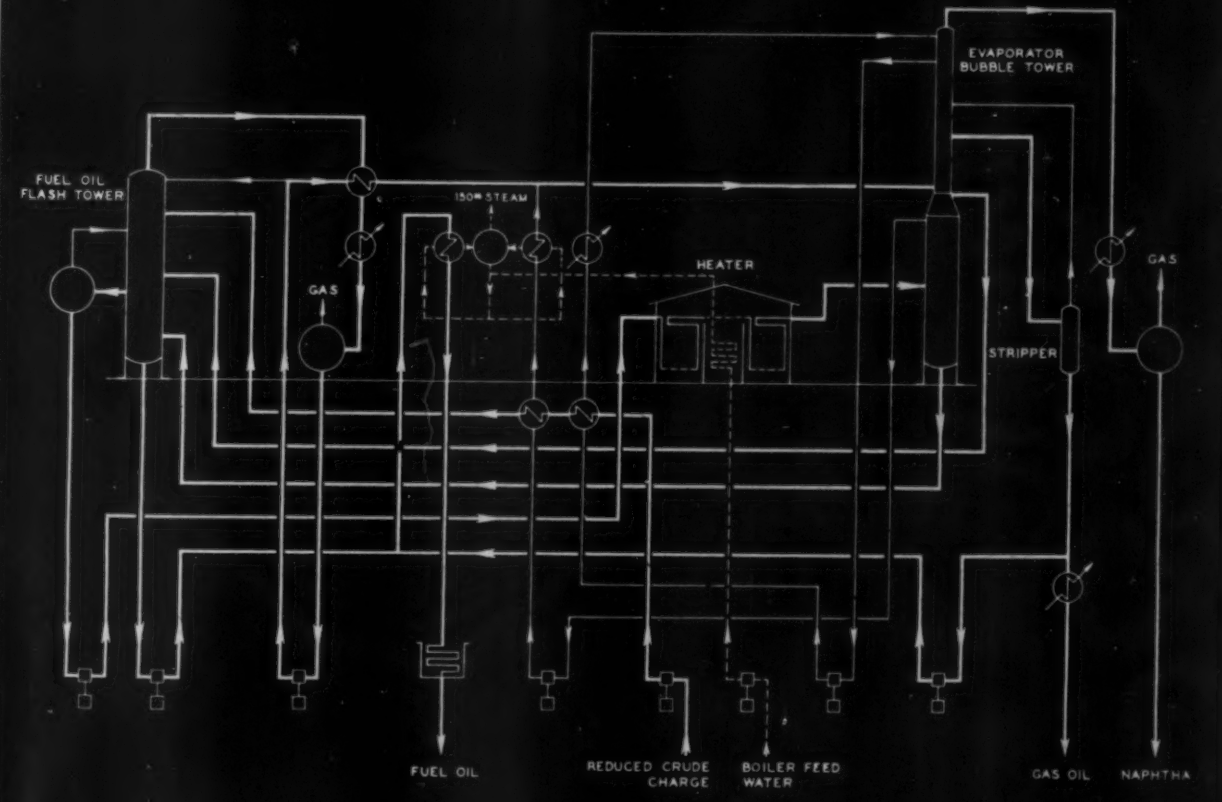
In the specialized field of reduced-crude processing, The Lummus Company offers several processes from which a selection may be made depending upon the refinery situation and marketing requirements. These processes are in general employed to increase gasoline production and reduce residual fuel oil production; they may be applied to furnish a gasoil feed for a combination or catalytic-cracking unit. Among the processes available for this purpose are Vacuum Reduction, Propane Deasphalting, Coking, and Viscosity Breaking. At the moment the urgency for reducing residual fuel oil productions is not great, as it was prior to the war because of the relatively large present demand for fuel oil and the general trend towards lower production resulting from the operation of extensive catalytic cracking capacity.

Lummus viscosity-breaking units (see flow sheet opposite) employ a number of novel features depending upon the requirements of the situation including two-coil cracking, vacuum reduction of the residue followed by cut-back with light cycle stock for the purpose of minimizing pour and viscosity, and flexibility in design which enables these units to process a wide variety of stocks efficiently.

Extensive thermal cracking facilities are now being built in one large plant by The Lummus Company, 420 Lexington Avenue, New York, which provided this discussion of the subject as well as the flow diagrams on the page opposite.



LUMMUS COMBINATION CRACKING UNIT



LUMMUS VISCOSITY BREAKER UNIT

COURTESY, THE LUMMUS COMPANY

REFINERY EQUIPMENT...

AIRETOOL CATALYST REMOVAL SYSTEM

For removing spent catalyst efficiently from vertical reactor tubes in reactor type towers of Polymerization units. The Airetool Catalyst Removal System illustrated above is effective and easy to operate.

Due to changing catalyst conditions, the method of advancing the cutter head upward into the catalyst is especially important. The cutter head is air operated while in contact with the catalyst. When the catalyst is dry and not extremely hard the cutter head works most effectively when advanced by the power feed method. When the catalyst is wet, sloppy or gummy it is only necessary to shut off the power and slip the hand cranks on the machine and the cleaning operation continues with a minimum of lost time.

Tubes are thoroughly cleaned in 3 to 7 minutes. The machine can be shifted from tube to tube quickly as it is equipped with swivel casters which permit easy handling and spotting under the tube.

Hydraulic pressure holds the housing of the machine securely against the bottom of the tube sheet of the tower and the catalyst cuttings drop by gravity and are conveyed into any suitable receptacle for disposal.

AIRETOOL
MANUFACTURING COMPANY
SPRINGFIELD, OHIO

AIRETOOL • • AIRETOOL • AIRETOOL • AIRETOOL

...engineered by AIRETOOL

Airetool Cleaners, Expanders and Refinery Specialties are designed to give long satisfactory service at minimum cost to the petroleum industry. Prompt service from the Airetool engineering department is available to oil refineries for advice on special tube cleaning, tube expanding and catalyst removal problems.

a. AIRETOOL EXPANDERS. Refineries use Airetool tube expanders for correct expanding and flaring of all types of tubes from ⅜" I.D. to 12" I.D. Illustrated is an Airetool G-100 complete with ironing-out attachment and thrust.

b. TUBE CLEANERS. Airetool manufactures tube cleaners for both straight and bent tubes ranging in size from ¼" to 24" I.D. To meet every refinery tube cleaning requirement Airetool provides various combinations of motors and cutters, brush and drill heads. Illustrated is an Airetool Tube Cleaner with a P-type head and high speed motor for use in straight tubes 2¼" I.D. to 4¾" I.D.

c. FORWARDLY SWING ARM HEAD for all bent tubes 1¾" to 12" I.D.

d. DOUBLE EXPANSION HEAD with front offset cone cutter for straight tubes 3" to 10" I.D.

e. AIRETOOL MOTORS. These more powerful motors feature slip-fit construction which is an important time saver as the motor can be taken apart and reassembled without special tools. Other features include the power seal by which the blades of the rotor are pressure held against the cylinder and balanced rotor which makes it possible to use a simple flat plate for thrust bearing surfaces. Proper oiling is assured by five points of lubrication at each bearing. Because of the power seal mentioned above the motor can be loaded down to 50 R.P.M. without stalling.

f. CC-450 DIRECT-DRIVE CONDENSER CLEANER. For tubes ⅜" to 1½" I.D. Other models, CC-375 and CC-325, Geared Drive. Available for tubes ⅜" to 2½" I.D.

ACCESSORY TOOLS. Airetool provides Calipers to accurately measure or check tube end I.D., thickness of headers, header walls, and valve bodies; Knock-out Tools, for removing tube stubs; a Cleaning Lance for Heat Exchanger Bundles; Parallel and Right Angle Gear Drives; Flycutters, and Combination Boring and Cut-off Tools.

Write for Catalog

A 1948 Oil Refinery Catalog which lists the principal products of the Airetool Manufacturing Company is now available. Write the Airetool Manufacturing Company, 352 S. Center Street, Springfield, Ohio.

AIRETOOL • AIRETOOL • AIRETOOL • AIRETOOL • AIRETOOL

Thermal Reforming Process

UNIVERSAL OIL PRODUCTS COMPANY

THE UOP Thermal Reforming Process has had such wide and successful application in all parts of the world that it ranks with thermal cracking as one of the best known processes in the petroleum refining industry. This process converts low-grade gasolines and naphthas into high octane gasoline by subjecting the charging stock to heat treatment under controlled time, temperature and pressure conditions. The process is flexible, not only in its ability to process various stocks, but also as to quality and quantity of products which can be made from these stocks.

Thermal reforming produces butane-butylenes and propane-propylene which can be polymerized to produce a high octane polymer gasoline which is an excellent blending material for improving the quality of other gasolines. The residuum produced by this operation has a high calorific value, and can be marketed as fuel oil.

As shown by the accompanying flow chart, a straight-run gasoline or naphtha fraction is charged to a heater wherein the reaction time, pressure, and discharge temperature are controlled to produce the desired products. The effluent from the heater passes through a pressure-control valve where the pressure is reduced. A quench stream from the fractionator is introduced after the control valve to reduce the temperature of the effluent and minimize coking.

The quenched products enter a combination flash chamber and fractionating column wherein the gas and gasoline are taken overhead, and the small amount of bottoms produced is removed from the lower portion of the vessel. A cracked naphtha side-cut from this column is used for quench and heat exchange duty. Naphtha side-cuts may be withdrawn if desired.

The overhead products from the fractionator are cooled, and the condensed, unstabilized gasoline is sent to a stabilizer to produce the desired vapor pressure. If polymerization of the gases evolved from the cracking operation is desired, an absorber for the receiver gases plus deep stabilization of the gasoline is employed to provide maximum recovery and utilization of polymerizable olefins available.

The bottoms withdrawn from the flash-chamber portion of the reformer column are further flashed in a residuum stripper. Light material removed from the residuum at this point is condensed and returned to the fractionating column, whereas the residuum is sent to storage for subsequent disposal as fuel oil. The properties of the residuum are regulated by suitable control of the reflux to the stripper.

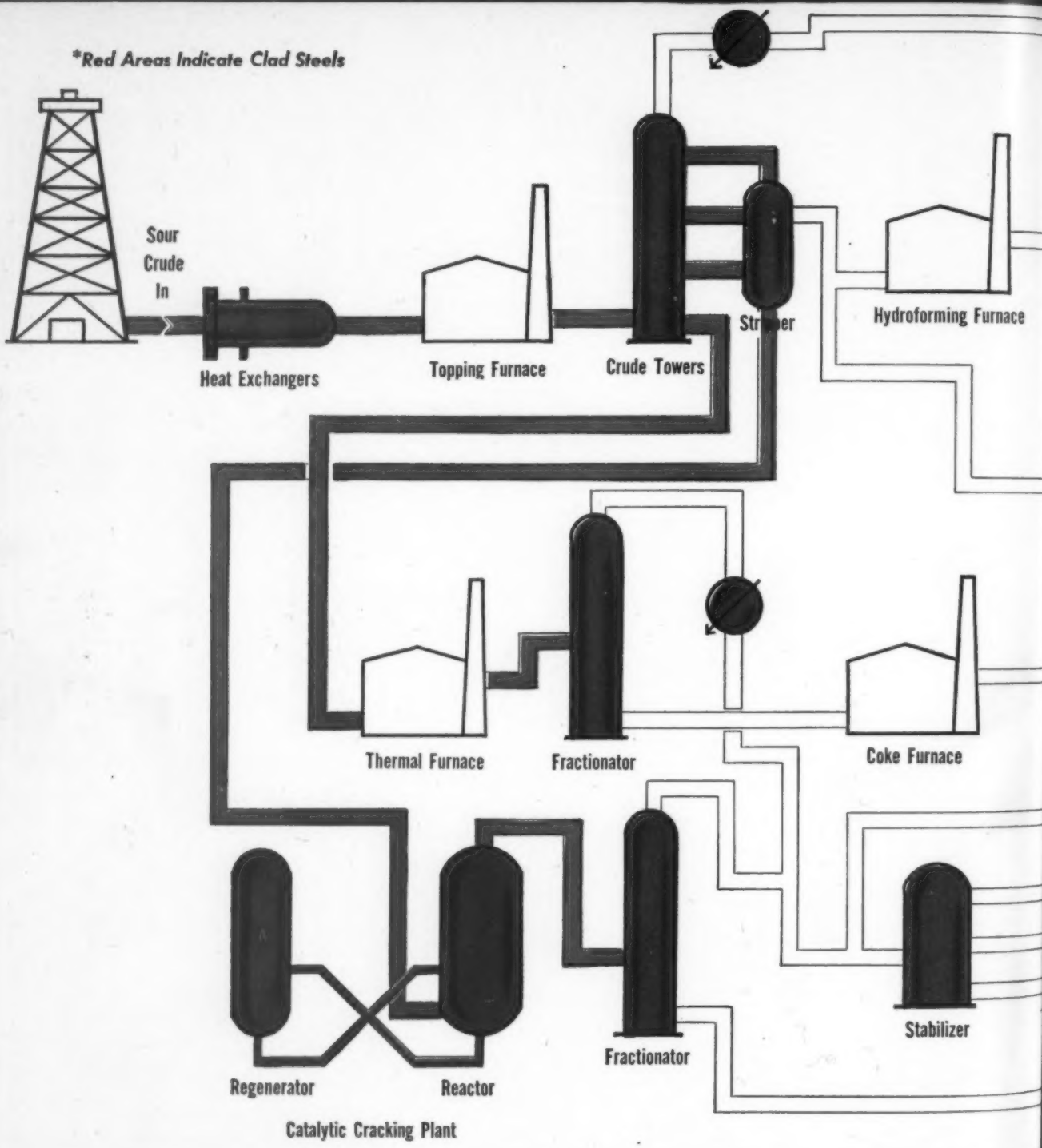
A typical result when reforming a 53° API Mid-Continent straight-run naphtha is shown in the following tabulation. It must be realized in considering these results that the yields and octane-rating improvement that can be derived from thermal reforming are dependent upon the characteristics of the charging stock. In order to design thermal reforming equipment for any refiner, his entire situation must be carefully analyzed and studied, not only from the standpoint of charging stocks available, but also from that of market requirements.

Thermal Reforming of Mid-Continent Naphtha

Charging Stock	
Gravity °API	53.0
Octane rating, ASTM (F-2) clear	40.5
Characterization factor	11.96
100 ml. distillation	
I.B.P. °F.	208
10%	263
50%	314
90%	367
E.P. °F.	389
Products	
Gasoline:	
Yield, vol. % of charge	81.3
Gravity, °API	56.1
Reid vapor pressure, lbs.	10.0
100 ml. distillation	
I.B.P. °F.	93
10%	145
50%	270
90%	360
E.P. °F.	398
Octane Ratings:	
ASTM (F-2), clear	70.2
+ 3 cc. TEL/Gal.	81.9
Research (F-1), clear	78.2
+ 3 cc. TEL/Gal.	91.3
Residuum	
Yield, Vol % of charge	4.2
Gravity, °API	18.2
Gas	
Yield, Wt. % of charge	15.3
Cu. Ft./Bbl. of charge	467

THERMAL REFORMING PROCESS
COURTESY, UNIVERSAL OIL PRODUCTS COMPANY
GAS
STABILIZER
STABILIZED DISTILLATE
GAS
RESIDUUM
GAS
FRACTIONATING COLUMN AND FLASH CHAMBER
RESIDUUM STRIPPER
CHARGE
HEATER

Applications for Clad Steels* in



LUKENS
Nickel-Clad Stainless-Clad
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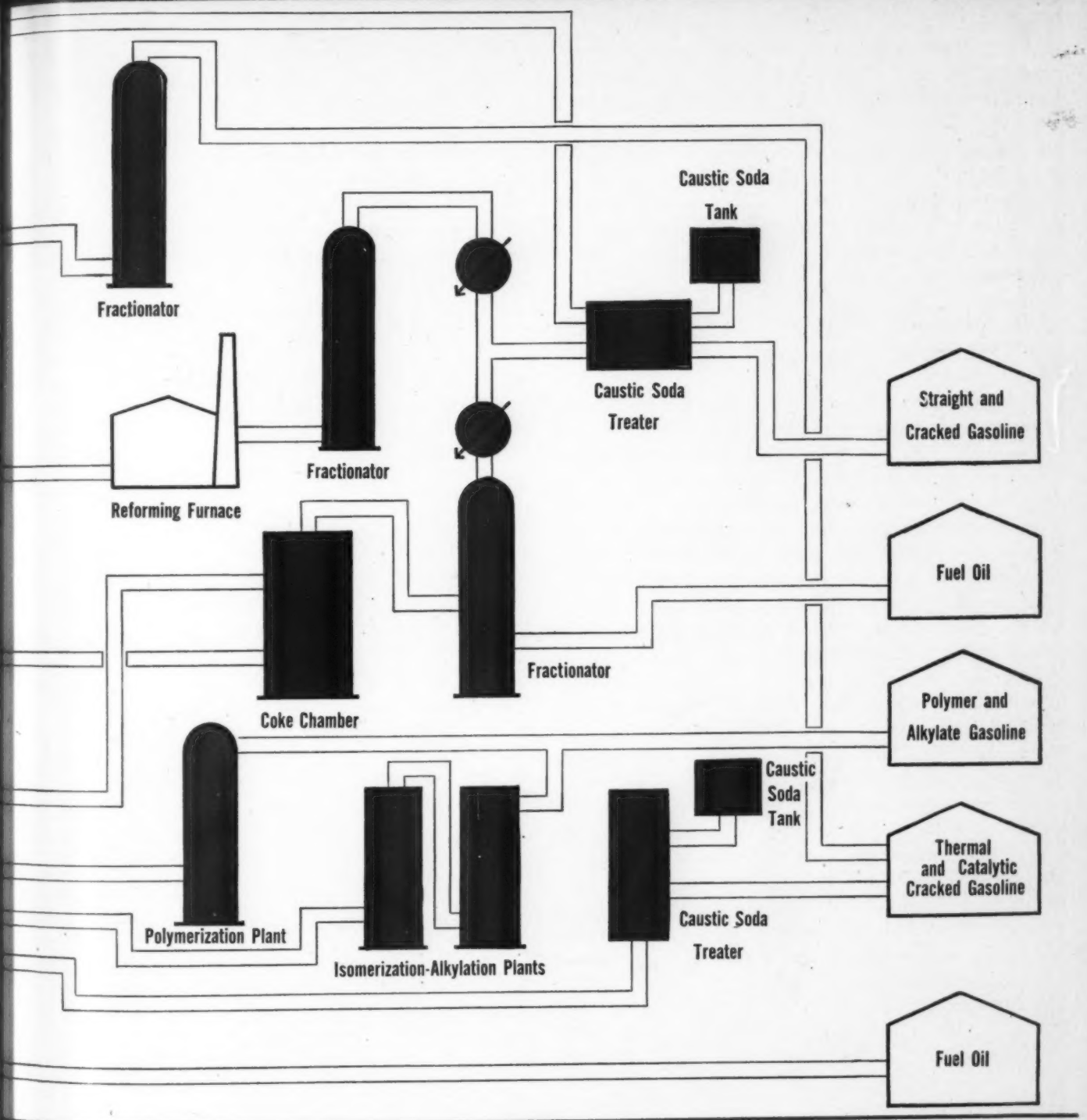
STEELS

SOLID METAL ADVANTAGES WITH CLAD STEEL ECONO

Refining Gasoline and Fuel Oil



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The new publications, No. 461 "Economical Corrosion Protection for the Petroleum Industry" and No. 399 "Fabrication of Lukens Clad Steels" will be sent on request to Lukens Steel Company, 404 Lukens Bldg., Coatesville, Pa.

Delayed Coking Process

THE M. W. KELLOGG COMPANY

DELAYED Coking is a continuous thermal-cracking process designed to produce clean (low Conradson carbon) catalytic-cracking feed stock from any type of reduced crude. A by-product of the process is a high grade of petroleum coke useful as a raw material in chemical industries.

The overall effect of the process on modern refinery operation is to increase the yields of motor gasoline and diesel fuel per barrel of crude oil processed, and to reduce the percentage of fuel oil.

There are at present about 13 Delayed Coking units operating or building in this country with a combined capacity of approximately 40 million barrels of reduced crude yearly.

Feed to the unit is usually a heavy reduced crude. After heat exchange with the hot products the feed is pumped to the base of a bubble-tower fractionator where any light components are flashed off. The bottoms of this bubble tower are pumped to a pipe furnace in which they are quickly heated to about 920° F. and discharged at a pressure of 10-70 psig into insulated coke drums to coke by their contained heat. The actual number of drums required is determined by the feed rate and carbon-forming characteristics of the reduced crude. At any given time half of the drums are coking and the other half are decoking. A specially designed transfer manifold utilizing an unusual switch valve makes it possible to feed any drum without coking liquid in lines or interrupting furnace operations.

A mixture of liquid and vapor enters the bottom of the coke drums but only vapor leaves the top, all liquid being converted in the drums to either coke or vapor. The vapors leaving the drums are fractionated in a bubble tower, producing gas and gasoline overhead and withdrawing light, medium and heavy gasoil as side streams. The heaviest gasoil, blended with the flashed residuum, is returned to the furnace.

The above description is applicable to recycle operation. However, Delayed Coking units are also operating on a single-pass basis.

Removal of coke from the drums is accomplished hydraulically by means of high impact-producing jets incorporated in special boring and cutting tools. The hydraulic method has been perfected to a point where the drums are cleaned speedily and safely and the coke transported to storage in a continuous mechanical operation.

Delayed Coking yields on a typical charging stock are given in the following table:

Table of Delayed Coking Yields and Product Inspection

Reduced Crude Charge	
Weight %	100
Gravity, °API	21.0
Conradson Carbon, %	6.0
Products	
1. **Stabilized Gasoline**	
Volume %	16.8
Gravity, °API	61.2
ASTM Distillation, °F.	
IBP	125
10%	175
50%	285
90%	365
EP	400
Reid Vapor Pressure, psi	4.0
CFRM Octane No.—Clear	60.0
2. **Gas Oil**	
Volume %	71.5
Gravity, °API	29.0
Conradson Carbon, %	0.03
3. **Dry Gas**	
Weight %	4.9
4. **Petroleum Coke**	
Weight %	12.9
Ash, %	0.15

The M. W. Kellogg Company, New York, has been appointed licensing agent for this process for the United States and foreign countries.

KELLOGG DELAYED COKING PROCESS
COURTESY THE M W KELLOGG COMPANY
CONDENSER
GAS
GAS SEPARATION DRUM
GASOLINE
COMBINATION STRIPPER
LIGHT GAS OIL
MEDIUM GAS OIL
HEAVY GAS OIL
FRACTIONATOR
FURNACE
COKE
COKE DRUMS
REDUCED CRUDE

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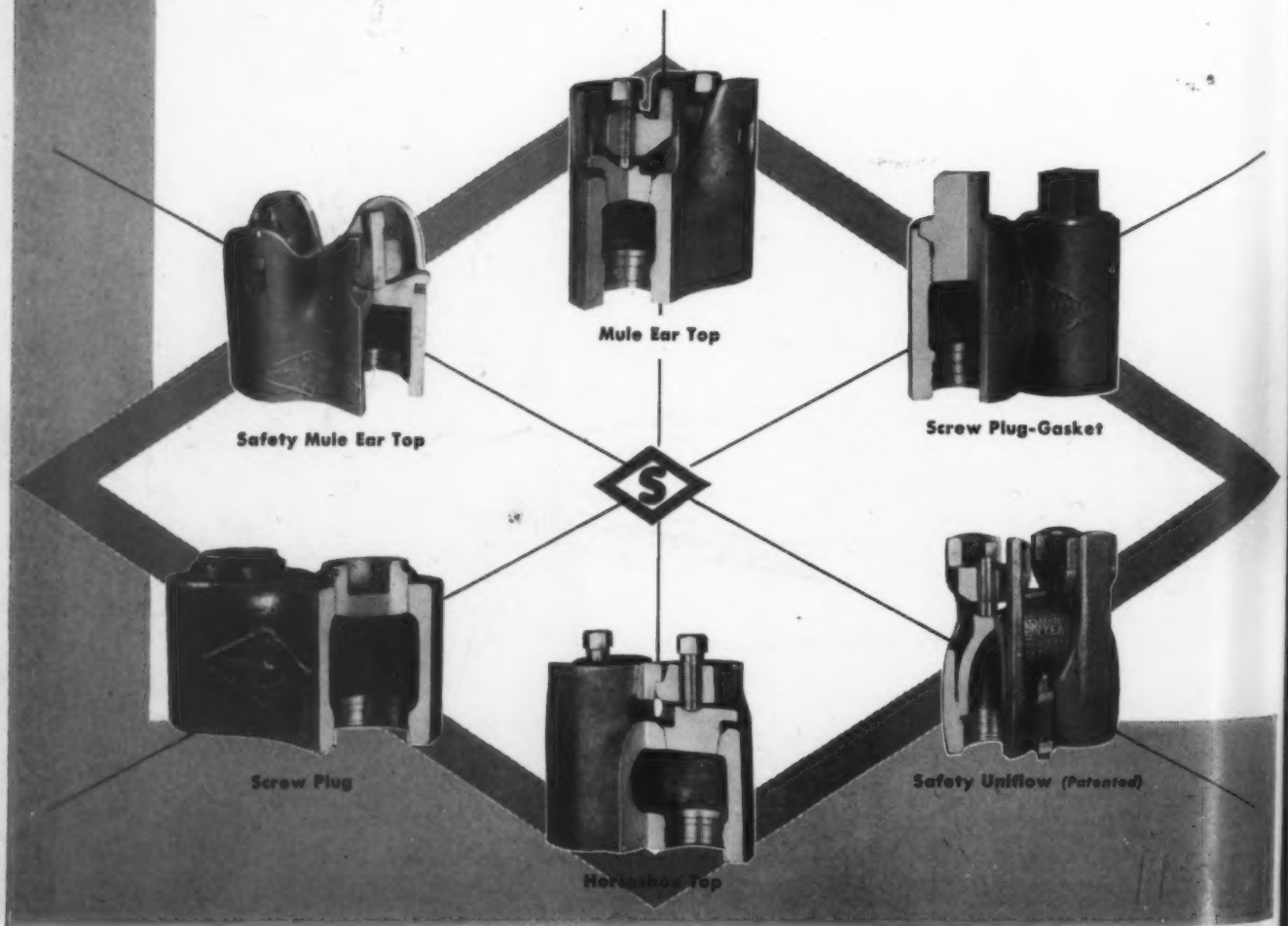

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Delayed Coking Process

FOSTER WHEELER CORPORATION

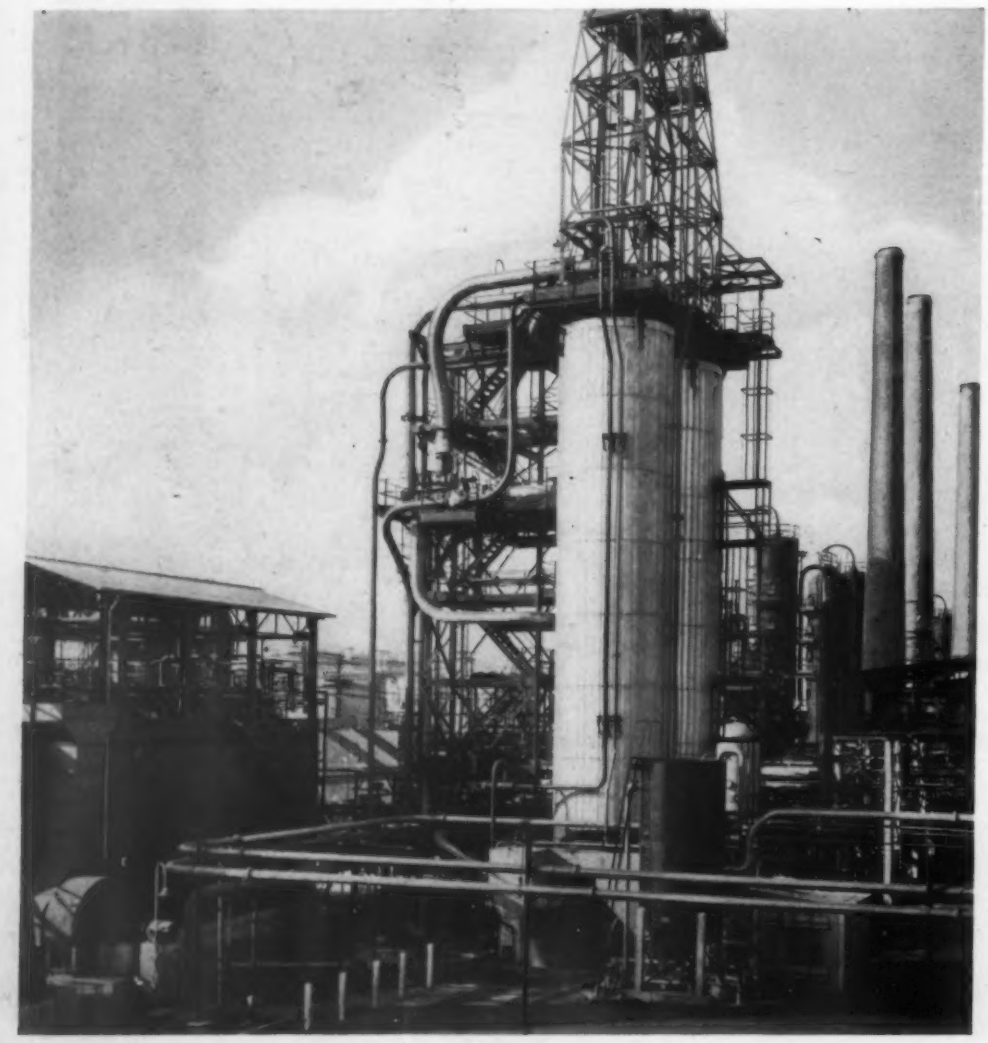
IN THE Thermal Delayed Coking Process stocks such as reduced crude, straight-run pitch or cracked tars are heated to a temperature sufficient to cause their heaviest components to turn to coke in a time chamber, from which the coke is intermittently removed.

The primary objective of delayed coking is to produce, from heavy carbonaceous oils, clean gasoils which are suitable charges for thermal or catalytic cracking. When demand and market realization for coke are adequate, delayed coking of the heavy surplus oil stock can be a very profitable operation. A yield as high as 20 percent of 400° F. end-point gasoline from the delayed coking of reduced crude represents an added asset in the overall economic balance.

Thermal delayed coking units may be designed to operate as viscosity-breakers when there is either no demand for coke at a fair return, or when there is a surplus supply of heavy residues.

Two or more coking chambers usually are provided for the intermittent removel of coke. The "hydraulic method" is looked upon with much favor by operating men because it consumes the smallest amount of time and produces the smallest percentage of coke fines. In this method a vertical hole about 18 inches in diameter is bored by a high-pressure water jet through the coke bed after the coke has been prepared by adequate steaming. The next step is the cutting of the coke into layers by using high-pressure water jets applied by means of rotating nozzles on the end of a pipe extending down to the bottom of the vertical hole. This cutting works upward until the entire body of coke is removed from the chamber. The water used for cutting the coke also may be used to convey the coke in a conveying system from the coke chamber to the stock pile.

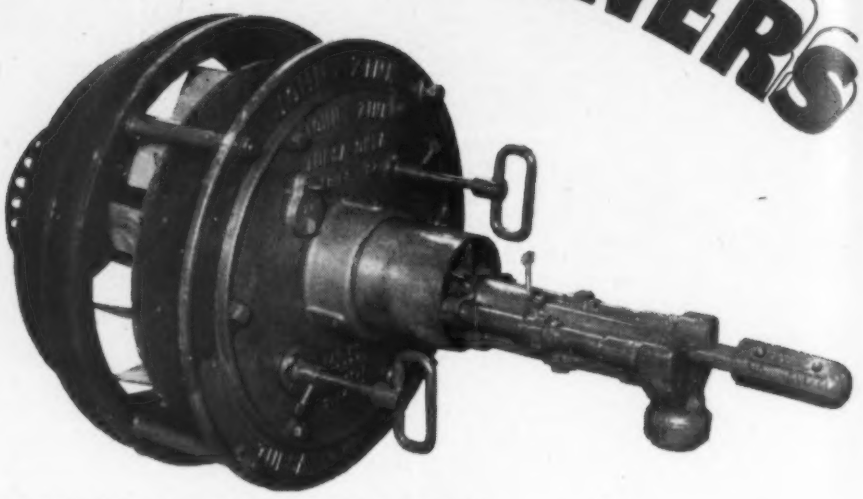
The charge, a heavy reduced crude or cracked tar, is pumped through the convection section of the "Downdraft" heater and enters the radiant section where it is heated to a temperature high enough to effect vaporization of all light and heavy gasoils contained in the charge. This liquid-vapor mixture enters the fractionating column and the unvaporized heavy fuel oil fraction of the charge is pumped from the bottom of this fractionating column into the other cell of the "Downdraft" heater where further vaporization takes place. This combination of heavy oil vapor and liquid residue (in which coke-forming substances are highly concentrated) enters one of the coking chambers where as a result of the combination of time and temperature, coke is formed. This coke accumulates until the chamber is filled and the stream is then diverted to the other coke chamber. The vapor leaving the coke chamber contains an appreciable quantity of low-boiling fractions, the result of cracking both in the heater coil and in the chamber itself. This vapor enters the fractionating column and is separated into gas and gasoline at the top of the column and a gasoil sidestream.



DELAYED COKING PROCESS
COURTESY, FOSTER WHEELER CORPORATION
Fuel Gas
Gasoline
Gas Oil
Fractionating Tower
Coke Chambers
Heater
Charge

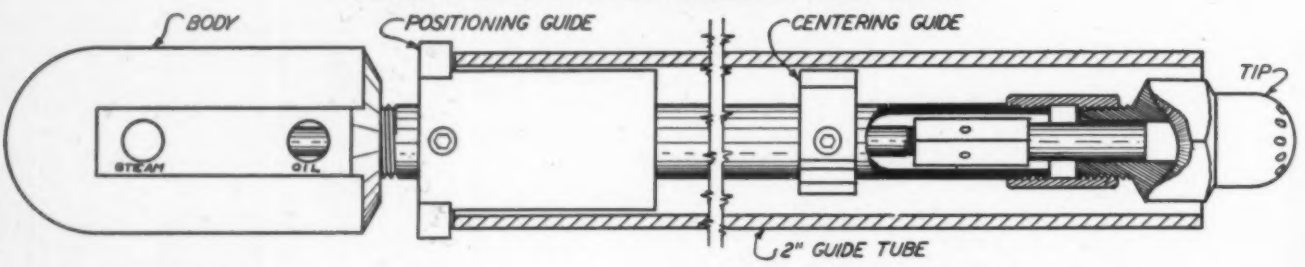
JOHN ZINK BURNERS

SERIES "Y" COMBINATION OIL and GAS BURNER



FUEL SCARCITY is no problem if you use John Zink Series Y Burners because this burner burns either, or both, gas and oil with the same heat pattern.
Features: Easy to remove fuel guns—feather weight type, one man can remove for inspection and cleaning while the boiler remains in service.
Easy to operate.
Simple in construction.
No tile necessary—may be installed in circles turned in firebrick.

JOHN ZINK STEAMIZER OIL GUN



HEAT PATTERN:

John Zink Streamizer Type Oil Gun produces a very stable, easily controlled fluffy yellow-white flame, the proportions of which can be suited to required applications.

TYPE OF FUEL:

Any non-corrosive liquid fuel heated so that its viscosity as delivered to the burner is not greater than 25 S.S.F. may be burned.

STEAM CONSUMPTION:

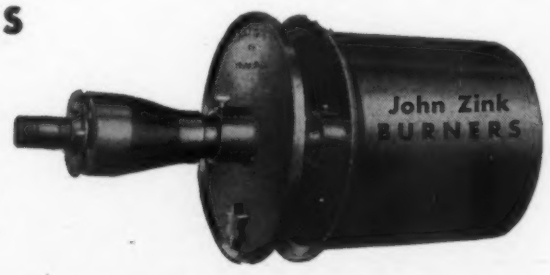
Metered tests on consumption of steam for atomization show that as little as 0.17 pounds of steam per pound of oil may prove satisfactory.

TURN-DOWN RANGE:

Wide turn-down range which requires no delicate adjustment is provided making the burner readily adaptable for automatic control. Capacities from 450,000 to 40,000,000 B.t.u./hr. are offered as standard. Higher capacities available for special applications.

HORIZONTAL BI-MIX BURNERS FEATURE:

Inexpensive to buy—easy to install. All in one piece—just brick into furnace wall.
Burns with a short flame, without the use of small ports, refinery gases, natural gas, oil vapors or a mixture of gases.
Slugs of water, oil or gasoline blow through them or run down over them without doing damage.
Can be used with high or low pressure gas.
Available for vertical or horizontal firing.



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Naphtha and Gasoil Polyforming Process

THE LUMMUS COMPANY

THE Polyform Process, licensed under patents of Phillips Petroleum Company and Gulf Oil Corporation, affords an efficient means for converting straight-run naphthas, gasoils and extraneous light hydrocarbons into high-quality motor gasoline. Polyform gasolines characterized by high research octane numbers possess excellent road performance at all speeds, and find use as the primary component of regular- and special-grade gasolines which must meet severe competitive specifications.

A Polyform plant is a self-contained unit, designed to process naphthas, gasoils and condensible light hydrocarbons simultaneously. The unit is complete with heaters, high-pressure fractionator, stabilizer and an absorber, which equipment affords the means to condense the light hydrocarbons for the furnace charge without necessity of refrigeration. Propane-propylene and butane-butylene produced in the process are largely condensed with normal cooling water supply and the balance is absorbed in the naphtha or gasoil feeds. Extraneous light hydrocarbons may be charged into the furnace-feed accumulator as a liquid or charged to the main high-pressure absorber. Condensed liquid gas is circulated with the feed and the presence of this material in the furnace charge enables substantially more severe temperature and pressure conditions than could otherwise be achieved. This, in addition to the polymerization of the light hydrocarbons, results in the high gasoline yield and octane associated with the Polyform process. By varying the charge rate of extraneous butane-butylenes, an efficient means of plant vapor pressure balancing is afforded.

The Polyform Process has found most extensive application in the processing of low-octane naphthas, although there are in operation a number of units charging gasoils. The process is particularly well suited for the conversion of refractory catalytic cycle stocks requiring more severe conditions than can be achieved in conventional thermal-cracking plants. Yields of 10-pound RVP gasoline ranging from 60-80 percent have been obtained from virgin gasoils, while yields of 50-70 percent will be derived from catalytic cycle stocks with ASTM octane of 72-76 for virgin stocks and somewhat higher for the cycle oils.

Naptha Polyform units charging a wide variety of heavy naphthas with and without extraneous condensibles achieve higher yields of higher octane gasoline than is possible with conventional reforming units operating in conjunction with catalytic-polymerization facilities.

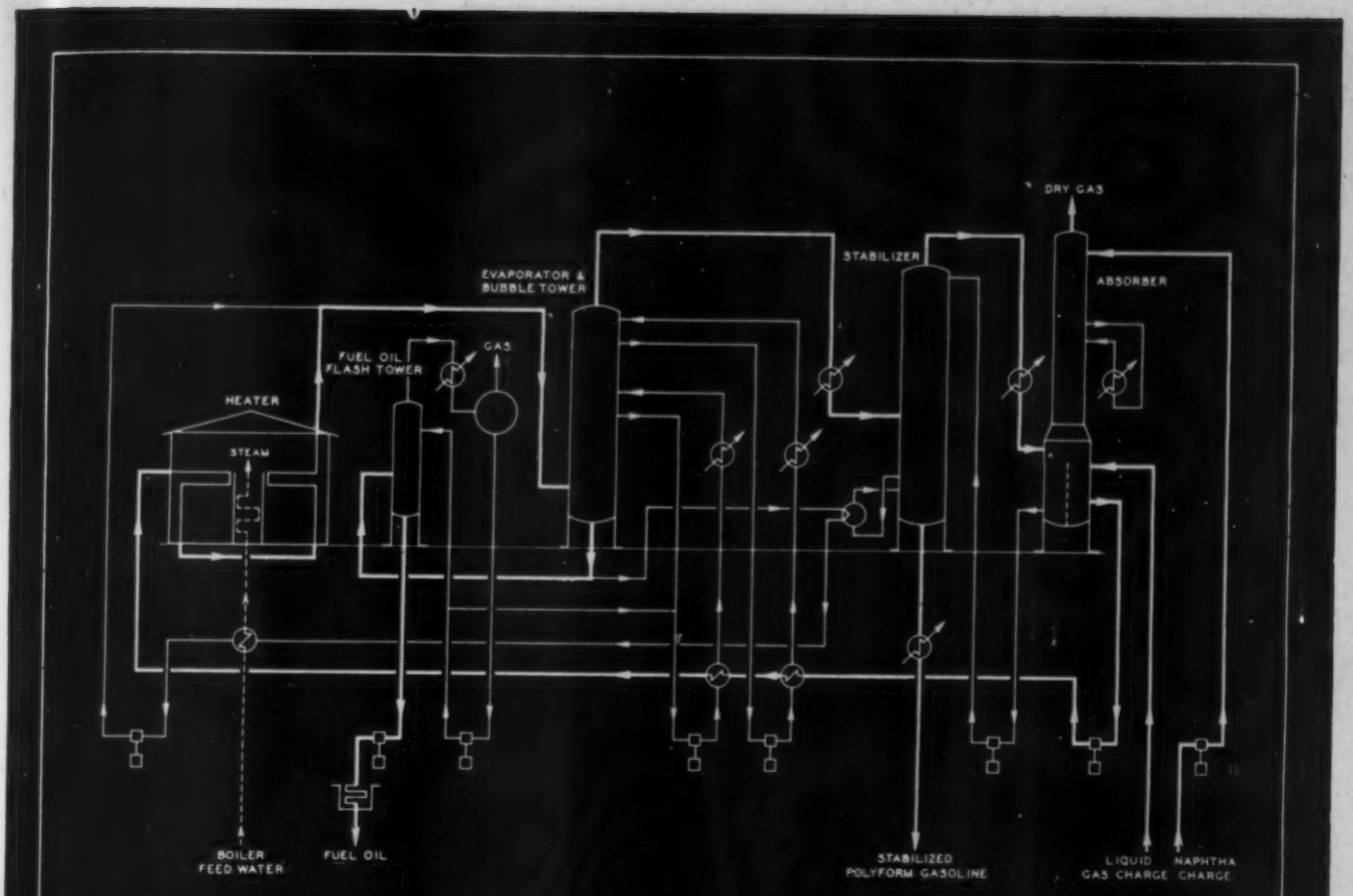
The flow sheets on the page opposite illustrate the process arrangement of typical naphtha and naphtha-gasoil polyform units. Either unit may accommodate outside gas charge as desired.

In the Naphtha Polyform unit the naphtha is charged to the absorber which is so operated that about 80-90 percent of the propane in the wet gas entering the tower is absorbed. At the base of the absorber the rich oil combines with stabilizer overhead condensate to make up the furnace feed which is charged with a high-pressure centrifugal pump to that the resultant transfer line pressure is 1200-1500 psi. Evaporator bottoms is employed to reboil the stabilizer and quench the transfer line to 650-700° F. The evaporator bubble tower operates at about 400 psi pressure. The net evaporator bottom flows to a fuel oil flash tower, the overhead of which is returned to the system, and the bottoms are delivered to fuel oil. The bubble tower is controlled with intermediate and top circulating reflux systems which preheat the furnace feed while the overhead flows through a partial condenser directly to the stabilizer, which tower operates at about 380 psig. The stabilizer may operate as a depropanizer if it is desired to produce butane-butylenes for alkylation, or may operate to produce specification-vapor-pressure gasoline with excess butane-butylene being taken overhead for conversion to gasoline. A large part of the propane-propylene and butane-butylene in the stabilizer overhead is condensed and flows to the base of the absorber, while the uncondensed wet gas flows through the absorber where it is stripped of propane and is then discharged as refinery fuel gas.

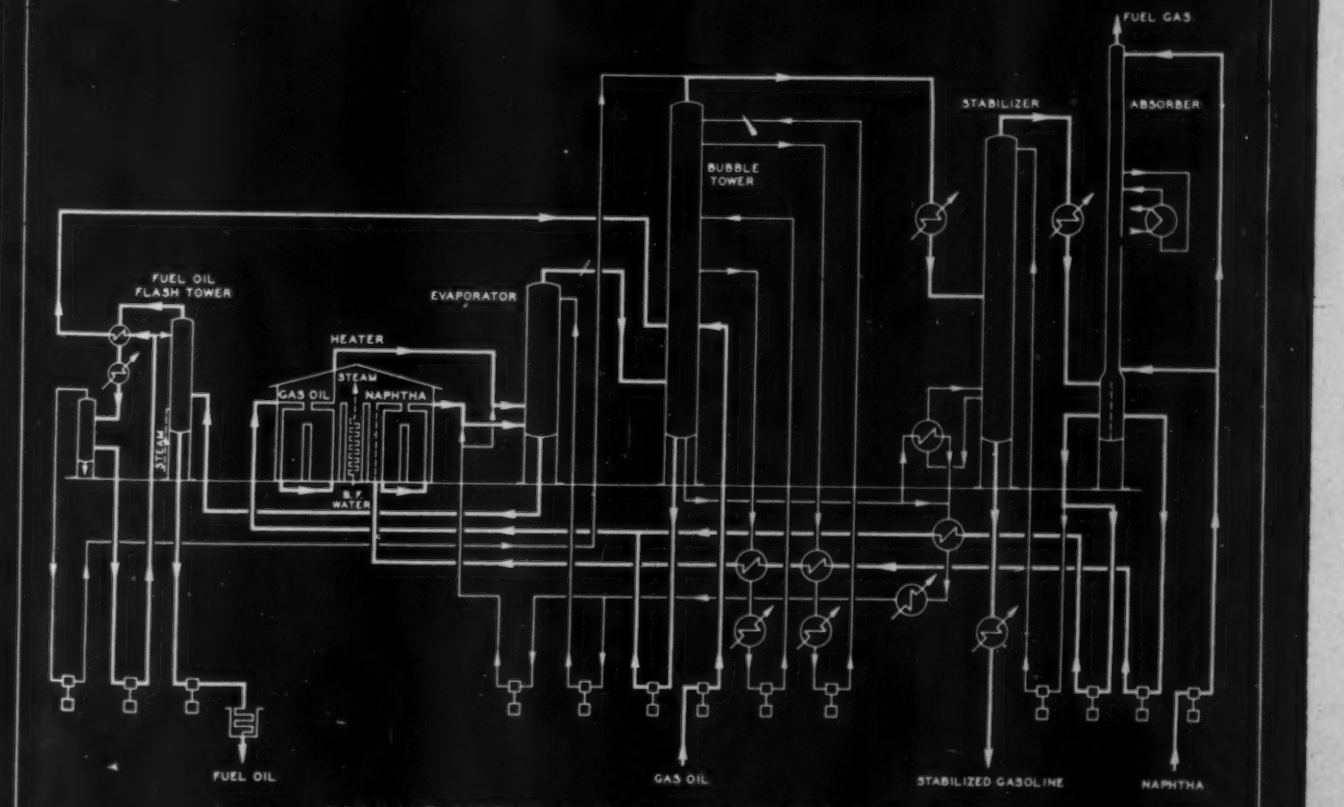
The Lummus floor-fired heater is ideally adapted for the exacting requirements of polyform service with separate heating and soaking sections. It has been found economical to use only radiant surface for the high-pressure oil service and generate steam in the convection section with forced circulation of boiler-feed water through an external steam drum. This not only brings about high fuel efficiency but provides sufficient steam for operating all of the high-pressure pumps and most of the low-pressure pumps so that the unit operates in substantially steam balance and requires no additional steam load from the refinery.

Polyform units are extremely simple and are typified by ease and stability of operation. Investment requirements for naphtha-polyform units are lower than for corresponding facilities for naphtha reforming plus catalytic or thermal polymerization.

The naphtha-gasoil polyform flow is substantially the same as for the naphtha polyform unit with only minor variations as required to accommodate the gasoil cracking.. In this plant the gasoil usually is charged to the bubble tower where it mixes with the gasoil recycle to form the furnace feed. High-pressure pumps handle the gasoil furnace charge and stabilizer overhead condensate, the latter being preheated against bubble-tower circulating reflux streams before mixing with the gasoil furnace charge downstream of the high-pressure charge pumps. Naphtha-gasoil units are typified by the same ease and simplicity of operation as in the naphtha-polyform units and the number of plants in service attest to the wide acceptance of the process.



LUMMUS NAPHTHA POLYFORM OR GAS REVERSION UNIT



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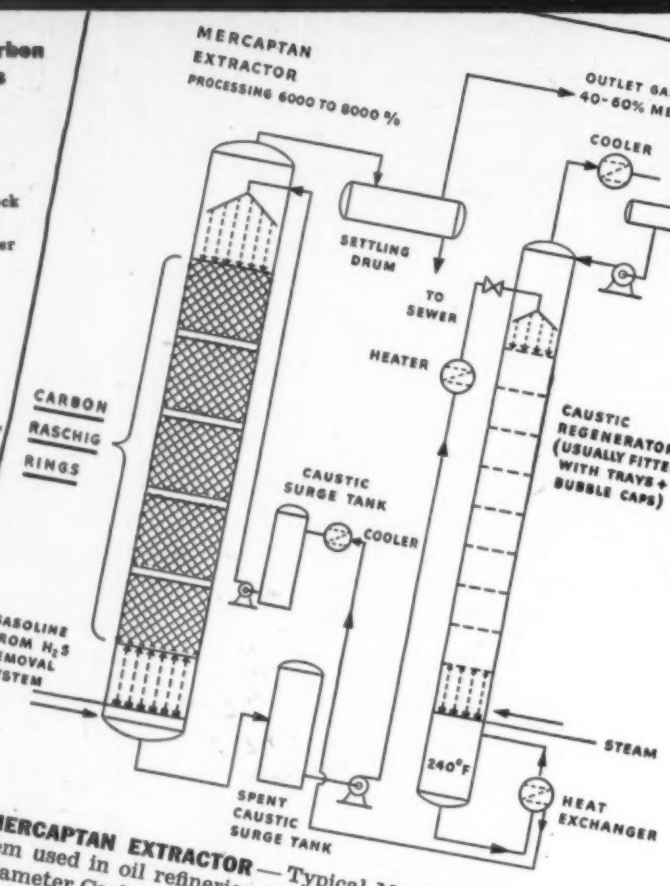
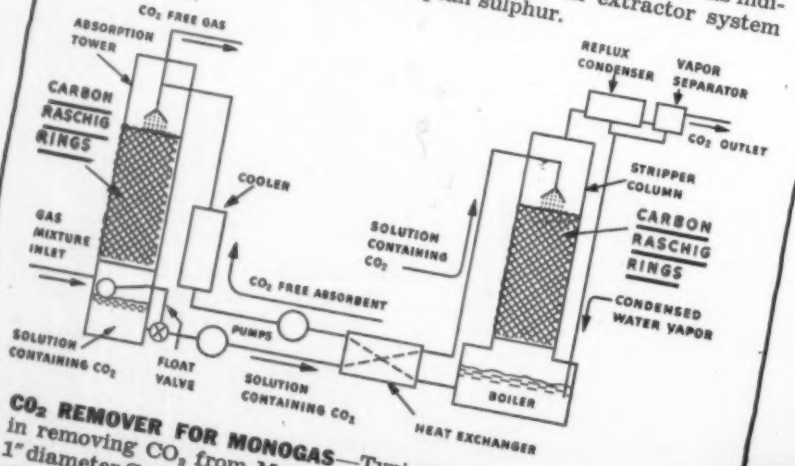


MERCAPTAN EXTRACTOR—Typical Mercaptan extraction system used in oil refineries. Extractor is usually packed with 1" diameter Carbon Raschig Rings, either solid or by trays as indicated. 10% to 20% NaOH solution used in extractor system removing up to 60% of Mercaptan sulphur.



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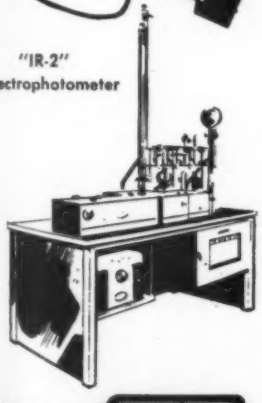

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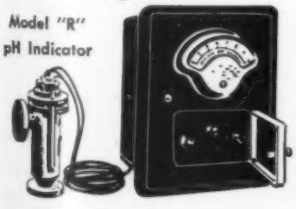

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Thermofor Pyrolytic Cracking Process

SOCONY-VACUUM OIL COMPANY, INC.

THE Thermofor Pyrolytic Cracking Process provides a practical method of converting hydrocarbons, ranging from ethane to reduced crude, into ethylene and aromatics. The TPC process was developed in the Socony-Vacuum Laboratories at Paulsboro, New Jersey and is based on techniques and knowledge acquired in the development of the TCC process. The process can be licensed from Socony-Vacuum Oil Company, Inc., and installed by authorized contractors such as Vulcan Copper and Supply Company and Rust Engineering Company.

The TPC system illustrated in the accompanying diagram comprises two main elements: a reactor and a heater. A bucket elevator is used to lift non-catalytic pebbles to the top of the unit where they are discharged into a heater. The heater serves to preheat the pebbles to the desired temperature level by direct combustion of fuel gas in the pebble bed. Any coke deposited on the pebbles in the cracking step is burned off in this stage. The hot pebbles flow by gravity from the heater into the reactor where the heat stored in the pebbles is utilized to preheat and crack the hydrocarbon feed. The cooled pebbles from the reactor flow through a throttling valve into the elevator which returns them to the top of the unit for reuse. A specially designed baffle system is used to prevent channeling of the pebble flow in both the reactor and heater.

The flow of hydrocarbons relative to pebbles through the reactor is countercurrent. The charge stock may be partially heated outside the reactor or may be introduced cold directly into the lower section of the reactor, where residual heat is available in the pebbles for preheating or vaporizing the feed. Extremely high rates of heat transfer, in the order of 15,000 BTU/hr/°F/cu. ft. of pebbles are obtained in this type of reactor. Thus the charge stock is preheated to high temperatures very rapidly and undesirable cracking at low temperatures is minimized. The cracking temperature is controlled by the temperature of the pebbles entering the reactor and the ratio of pebble circulation rate to hydrocarbon charge rate. The hydrocarbon contact time is controlled by the vapor rate through the reactor and the bed depth. The cracked vapors leaving reactor are quenched immediately to prevent decomposition of the ethylene.

The TPC process is extremely flexible with respect to the type of charge stock processed. It can be designed to crack either gaseous or liquid feedstocks. Crude oils or reduced crudes can be handled just as easily as distillate stocks since any carbon formed on the pebbles during cracking is continuously removed from the reactor. Carbon deposited in this manner is an advantage with the TPC process since it lowers the external fuel requirements in the heating stage.

Charge stocks ranging from ethane to reduced crudes have been evaluated for ethylene and aromatics production. The preferred operating conditions vary with the charge stock but generally fall within the following range:

Pebble to hydrocarbon feed ratio, weight	10-20
Pebble preheat temperature, °F.	1400-1800
Hydrocarbon contact time, seconds	0.1-2.5
Pressure, psig.	0-5
Steam, percent weight of charge	0-100

The product distribution obtained from TPC cracking East Texas Crude for ethylene is shown in Table I. The aromatic content of the 150-500° F. liquid fraction from an operation such as that shown in Table I would be about 50 percent. The aromaticity can be increased to 90-95 percent by increasing the contact time. Typical yields of pure aromatics from this type of operation on East Texas Crude are as follows:

Aromatic	Yield—percent wt. of charge
Benzene	7.8
Toluene	3.6
Xylenes	1.2
Other C_8 Aromatics	0.7
C_9 and C_{10} Aromatics	1.0
Naphthalene	2.2
1—Methyl Naphthalene	1.0
2—Methyl Naphthalene	0.8
Total	18.3
Higher boiling liquid	20.0
Ethylene	21.6
Ethylene Mol percent of C_2 and lighter	31.9

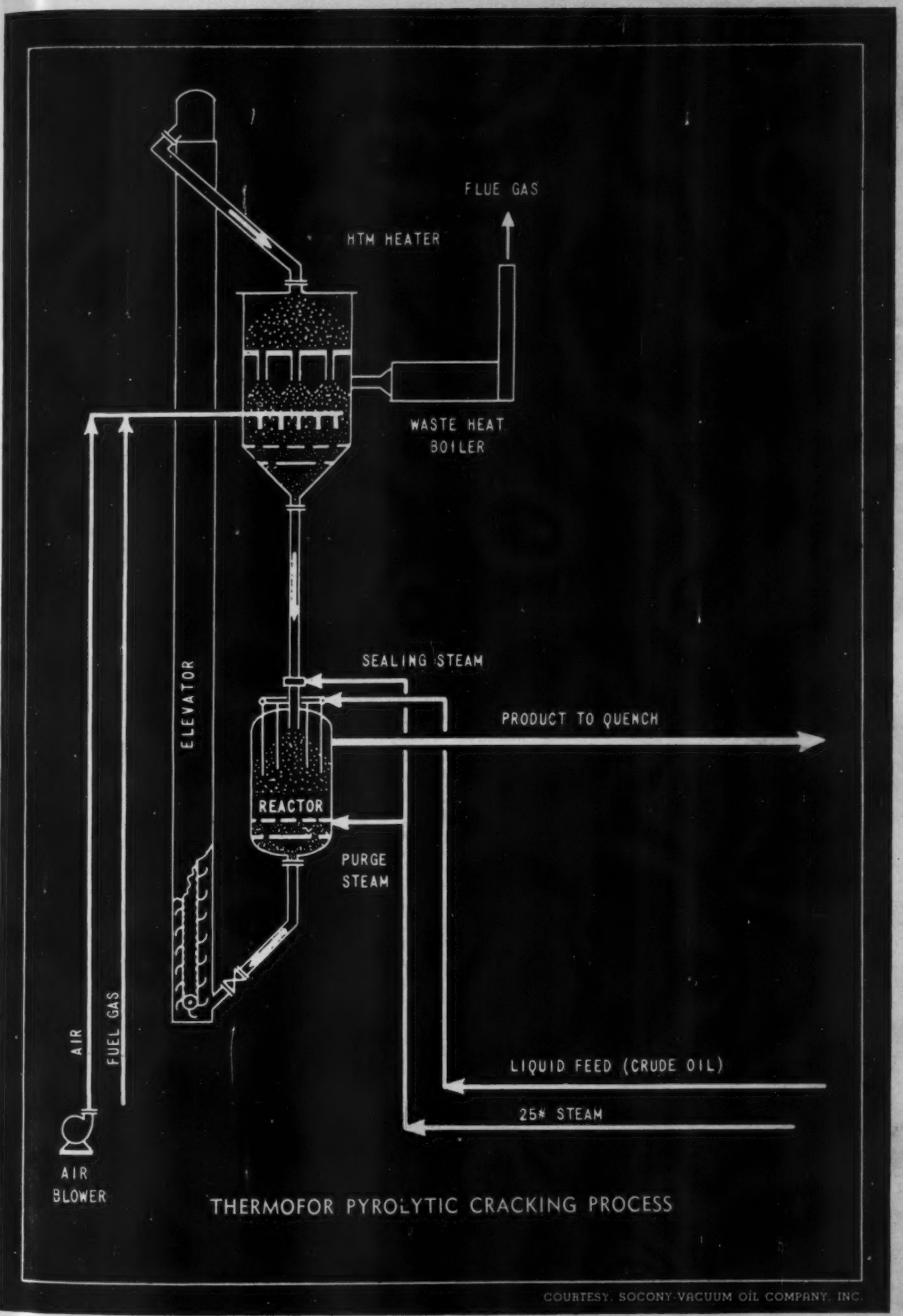
The application of the TPC process is not limited to the production of ethylene and aromatics. The cracking of liquid petroleum stocks to produce gaseous fuels for use either directly or as enrichment for industrial or domestic heating is potentially also very attractive.

Table I

TPC Cracking East Texas Crude for Ethylene Production

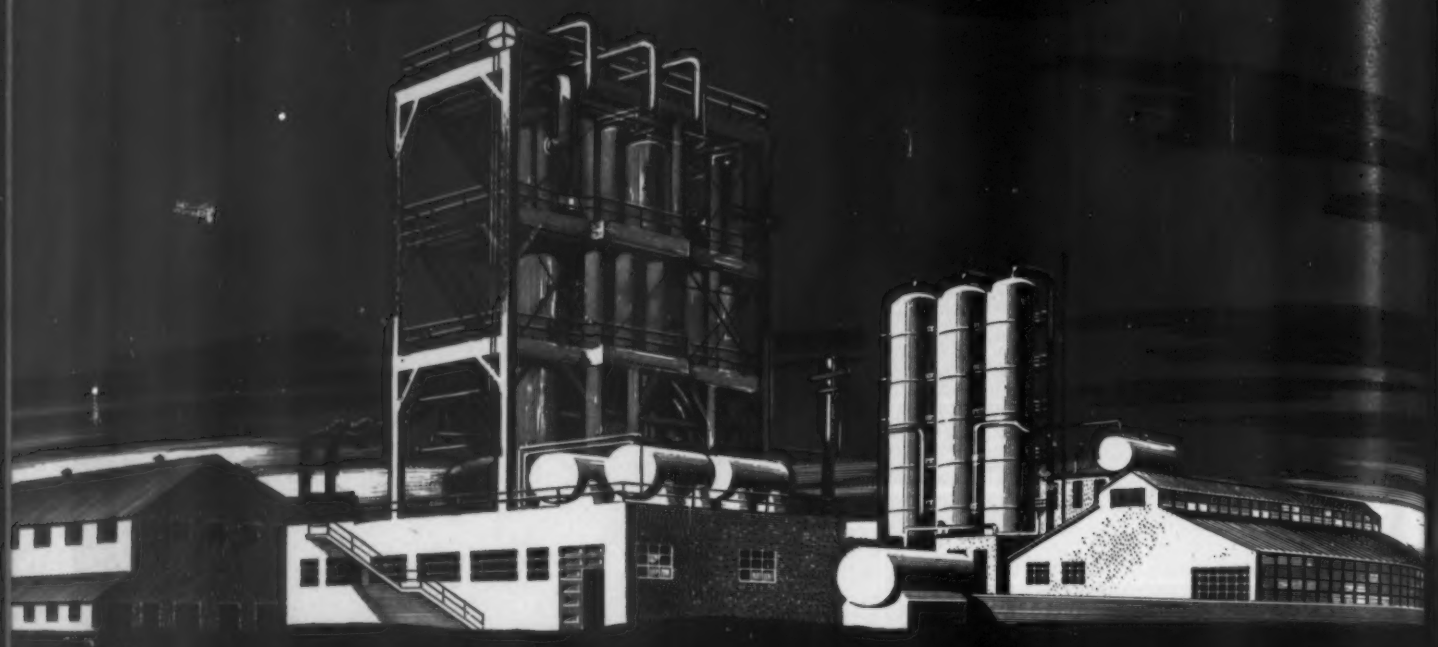
Properties of Charge Stock:	
Gravity, A.P.I.	38.4
Aniline Point, °F	172
Boiling Range, °F	
IBP	148
10 percent	290
50 percent	595
80 percent	980
Pebble Preheat Temperature, °F.	1515

Aromatic	Yield—percent wt. of charge
Hydrogen	0.7
Methane	14.1
Ethylene	22.3
Ethane	4.3
Propylene	12.2
Propane	0.7
Butadiene	3.1
Butylenes	3.0
Butanes	0.6
Total Liquid	37.2
Coke	1.8
Total	100.0
Mol percent ethylene in C_2 and lighter gas	36.7



THERMOFOR PYROLYTIC CRACKING PROCESS

COURTESY, SOCONY-VACUUM OIL COMPANY, INC.



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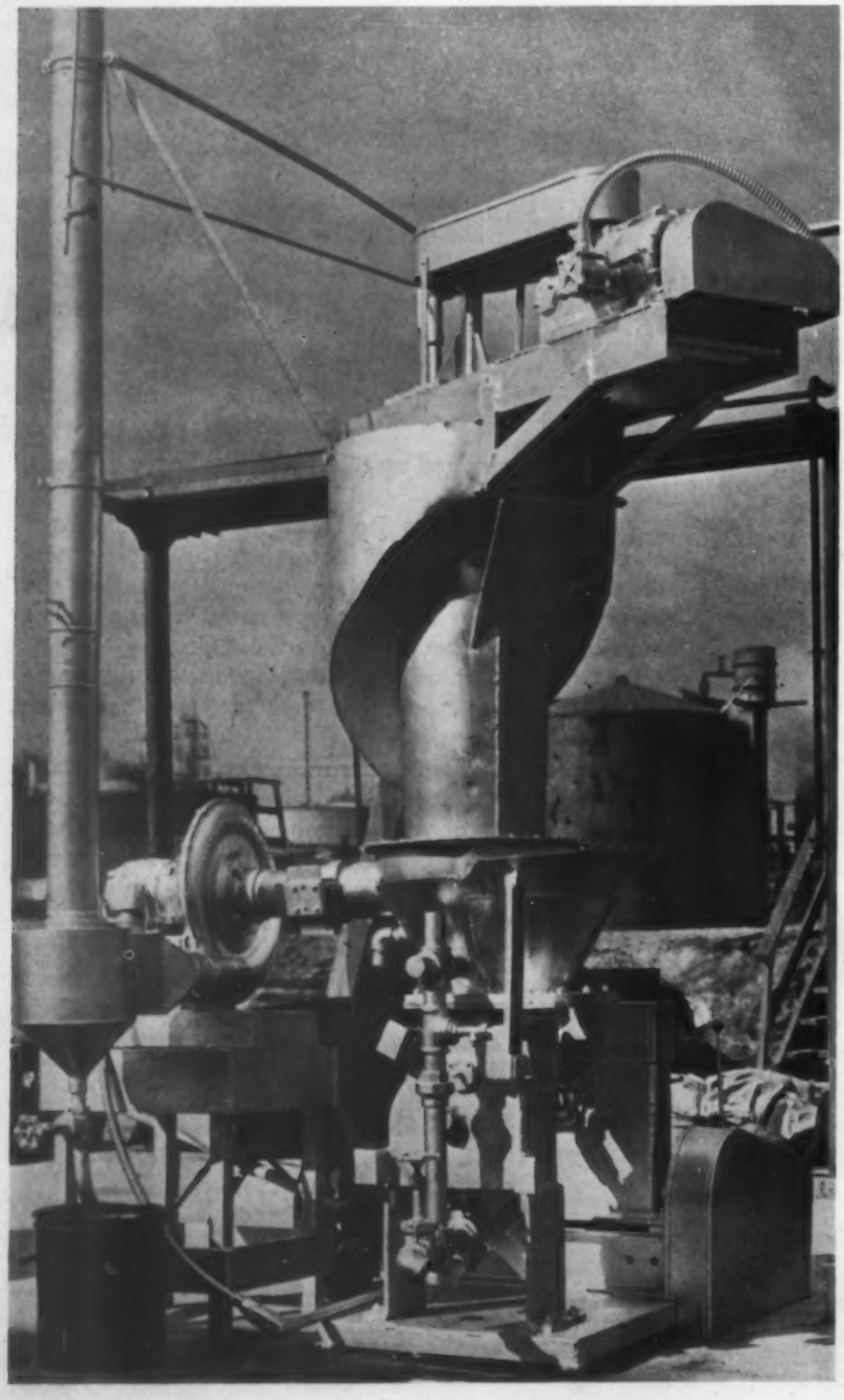
Continuous Shale Retorting Process

UNION OIL COMPANY OF CALIFORNIA

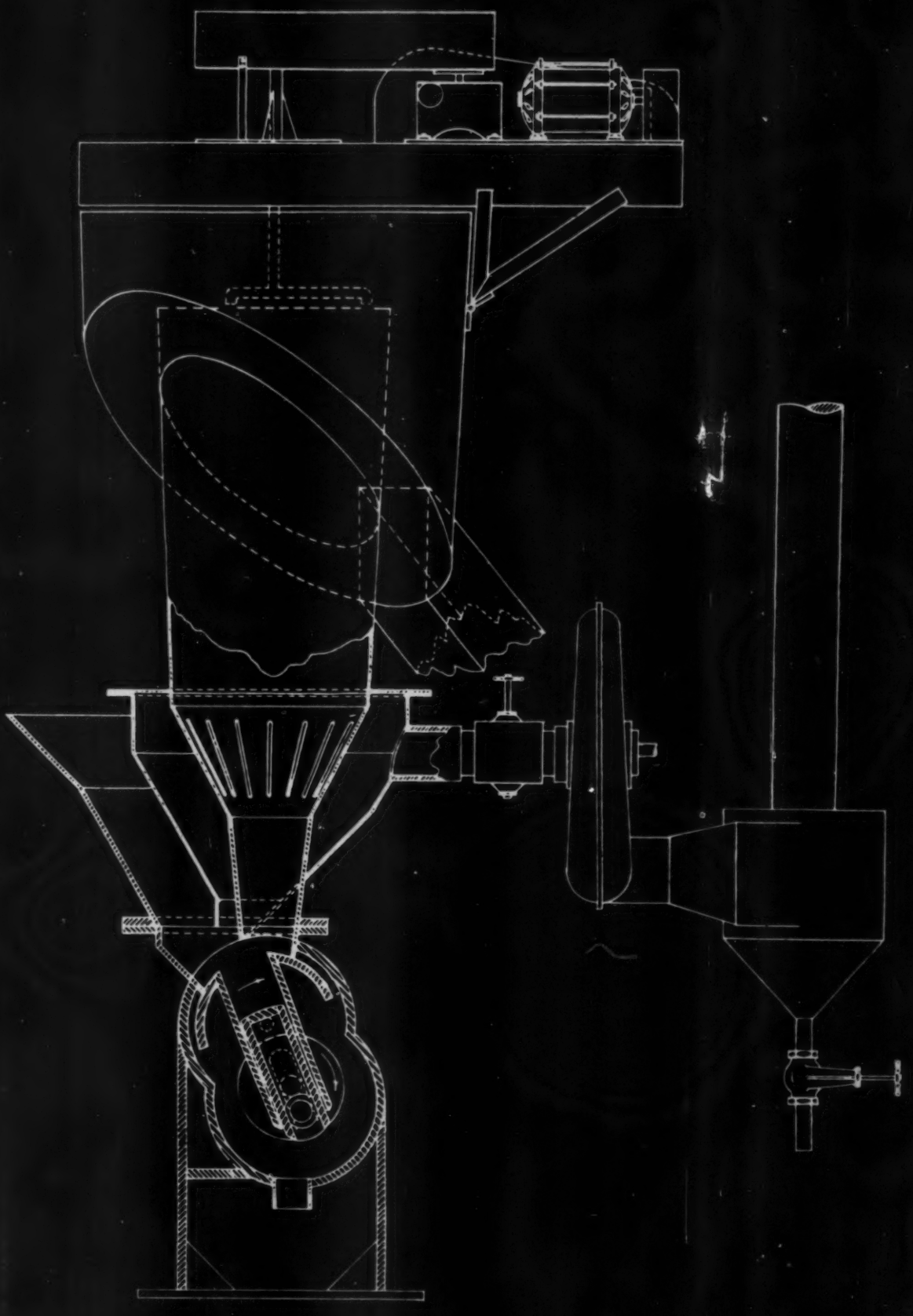
THIS continuous shale retorting process employs a new underfeed principle with countercurrent flow of shale and air. The pilot plant shown in the accompanying photograph has a shale throughput capacity of about 2 tons per day.



Shale, introduced to the kiln at the bottom, passes through a liquid seal and drops into the feeder mechanism by gravity. Subsequently it is forced upward through a slotted cone and above that through the slowly expanding shell of the kiln itself. A blower takes suction from the space outside the slotted cone and above the liquid seal and draws air down through the moving body of shale in the kiln. This air is first preheated by the hot spent clinker at the top of the kiln, burns off the residual carbon on the clinker in the next portion of the bed to supply the heat for distillation of the oil from the shale in the central section of the kiln. In the lower portion of the bed the distilled oil together with the combustion gases are cooled to nearly atmospheric temperature by heat exchange with the cold shale being fed upward. The oil and gaseous products are disengaged from the shale bed through the slotted cone and are drawn through a cyclone where the oil is separated and the gases are taken overhead. The spent clinker at the top of the kiln is broken up by a rotating plow and discharged into an annular collecting chute leading to a disposal bunker.

All the heat necessary for retorting the shale is supplied by combustion of the residual carbon on the clinker, and the heat available from burning the stack gases is more than sufficient to meet the power requirements of the plant. Since all heat exchange occurs in the kiln itself, expensive interchangers are eliminated and no cooling water is required.

This process, being continuous and countercurrent, has many advantages over shale retorting methods presently used. Nearly perfect heat exchange is attained in the unit, allowing the use of a very short bed. Recovery of oil from the shale consistently averages 100 per cent or better of the standard assay value for the shale being fed. The unit has a very high throughput for its size, and the equipment required is comparatively simple and inexpensive.



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Catalytic Oil Conversion Processes

THE large scale development and adoption of catalytic cracking processes is one of the most impressive and dramatic achievements of the process division of the oil industry.

Growing demand for gasolines of higher anti-knock properties brought about the design and construction of the fixed bed catalytic cracking units in the middle and late 1930's. These units were intended for peacetime operation in competition with the thermal cracking processes, and were successful, several plants being placed in commercial operation. The good yields of high quality gasoline obtained through catalytic cracking stimulated intensive research and development activity on the part of several major operating and engineering companies. Continuous catalytic cracking processes were devised and pilot plants built.

War brought a heavy demand for aviation base stock and the industry initiated a tremendous construction program to create the needed catalytic cracking capacity. All types of catalytic cracking processes were included in this program. Plants were built with fixed bed reactors, with moving bed reactors and bucket elevators for catalyst transfer between separate reaction and regeneration vessels, and with "fluid bed" reactors and regenerators having catalyst transfer by vapor carrier system. The plants, numerous and large, based on little operating experience, except in the case of the fixed bed process, were vitally essential to our war effort, and were to a remarkable degree successful from the time of initial operation.

This wartime program in a brief period established catalytic cracking as the predominate process for the conversion of gasoils into gasoline, and relegated thermal cracking to a secondary position.

The catalytic cracking of naphthas for octane improvement has found considerable application in recent years. Hydroforming, a process in which naphthas are reacted in the presence of hydrogen, was originally developed as a means for converting straight run naphthas to high octane gasoline levels. The process gives high yields of aromatic materials and as a consequence was extensively employed for the production of toluene for explosives manufacture during the last war. Installations at comparatively few refineries proved adequate to supply the bulk of our wartime needs. The units built then for toluene production continue to serve now in gasoline improvement operations.

Dehydrogenation processes found considerable wartime application in the manufacture of butylenes from butanes, but such plants have in many cases been shut down now that the critical shortage of butylenes is past.

Hydrogenation processes, while capable of a number of useful refinery applications, have found their major employment in the conversion of isooctenes from catalytic polymerization units into isooctanes for aviation fuels.

Thermofor Catalytic Cracking Process

HOUDRY PROCESS CORPORATION

THE Thermofor Catalytic Cracking (TCC) Process is of the moving-bed type and is licensed by Houdry Process Corporation.

As shown in the flow diagram the major parts of the catalytic section consist of reactor, kiln, and catalyst elevators. The reactor is a cylindrical vessel with catalyst and vapor inlets at the top and a catalyst-vapor disengager and oil-purge section at the bottom. Catalyst pellets flow downward by gravity, along with the hydrocarbon vapors, forming a solid moving bed. The depth of catalyst bed in the reactor is variable and controls the space velocity. After passing through the reaction zone the catalyst contains a small amount of carbonaceous deposit. Oil vapors and adsorbed oil are purged from the catalyst and surrounding voids in the purge zone after which the catalyst flows into the spent-catalyst elevator which operates at atmospheric pressure. This elevator transports the catalyst to the top of the kiln where it is discharged into a hopper feeding into the kiln. The catalyst flows downward by gravity through the kiln and the carbonaceous deposit is removed by zone regeneration with air at about 1.0 pound gauge pressure. Each burning zone has its own air supply and water-cooling coils, permitting a close control of catalyst temperature which thereby protects the catalyst from deactivation by excessive temperatures.

The kiln is designed to effect substantially complete removal of carbon from the catalyst by providing the required number of zones and residence time in each zone. Regenerated catalysts from commercial kilns normally average 0.2 weight-percent carbon or less. Most of the heat of combustion is converted into high-pressure steam by circulating water through the cooling coils and flashing in a separate steam drum.

The regenerated catalyst is returned from the bottom of the kiln by a second elevator to the reactor hopper, completing the catalyst cycle. On units charging less than 7000 barrels per day, a single elevator having partitioned buckets carries both spent and regenerated catalyst streams.

It should be noted that the elevators operate at atmospheric pressure, providing a positive separation of the kiln and reactor and assuring that a momentary upset on either kiln or reactor will not force a shutdown of the unit.

Catalyst within the TCC unit is maintained at maximum particle size by continuously by-passing a small portion of the regenerated catalyst stream through an elutriator where fines are separated with countercurrent air flow. Whole catalyst pellets from the elutriator then flow by gravity to the regenerated catalyst elevator.

An inert-gas seal is used to prevent the loss of hydrocarbon vapors out the top of the reactor. As shown, flue gas from the kiln or from the flue-gas generator is introduced at the bottom of a catalyst seal leg located between the feed hopper and reactor. The length of this seal leg is sufficient to provide a pressure drop exceeding the reactor pressure. The main part of the seal gas is vented to the atmosphere from the feed hopper, with a small portion passing through the reactor.

Aside from the catalytic equipment described above, heat exchange, fractionation, etc., are conventional.

The scheme of charge preparation shown in the flow diagram consists of heating reduced crude and flashing at approximately 15 psig pressure. The flashed distillate comprises the TCC charge. If desired the flashed residue may be vacuum-distilled or propane-deasphalted to obtain additional charge. In this situation the recovered heavy gasoil is heated to 700-800° F. and injected as a liquid at the top of the reactor.

The following table presents typical yields from a commercial plant using synthetic catalyst and cracking a coastal naphthenic charge stock in an essentially single-pass operation:

Cracking Coastal Naphthenic Gasoil for Motor Gasoline

	Vol. Percent	°API	Aniline Point °F.	DISTILLATION 10 Percent	DISTILLATION 50 Percent	DISTILLATION 90 Percent	OCTANES CFR-M Clear	OCTANES CFR-M 3 cc TEL	OCTANES CFR-R Clear	OCTANES CFR-R 3 cc TEL	Percent S
Charge	100	28.1	136	472	528	624	...	...	...	...	0.14
Products:											
10lb R.V.P. TCC Motor-Gasoline*	53.9	53.3	...	120	240	374	83.5	90.3	94.2	100.0	0.019
Excess C_4 Fraction	12.1	...	...	...	...	...	...	...	...	...	...
Catalytic Gasoil	34.0	21.7	88	470	510	600	...	...	...	...	0.07
Liquid Recovery (C_4+)	100.0	...	...	...	...	...	...	...	...	...	...
Dry Gas, Weight Percent	7.6	...	...	...	...	...	...	...	...	...	...
Coke, Weight Percent	2.1	...	...	...	...	...	...	...	...	...	...

SUMMARY OF OPERATING CONDITIONS

Total Space Velocity V./Hr./V.**	0.75	Vapor Inlet Temp. °F.	924
Catalyst	Bead	Catalyst Inlet Temp. °F.	950
Catalyst Activity Index	36	Avg. Reactor Pressure, psig	5.0
Process Steam	None		
Type Flow	Concurrent		

* Cu Dish Gum—6 mg; ASTM Gum—3 mg; Oxygen Bomb Stability 20 + Hrs.
** Contains 5.3 vol. percent of internal recycle.

THERMOFOR CATALYTIC CRACKING PROCESS
COURTESY, HOUDRY PROCESS CORPORATION
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POUNDS PER HOUR
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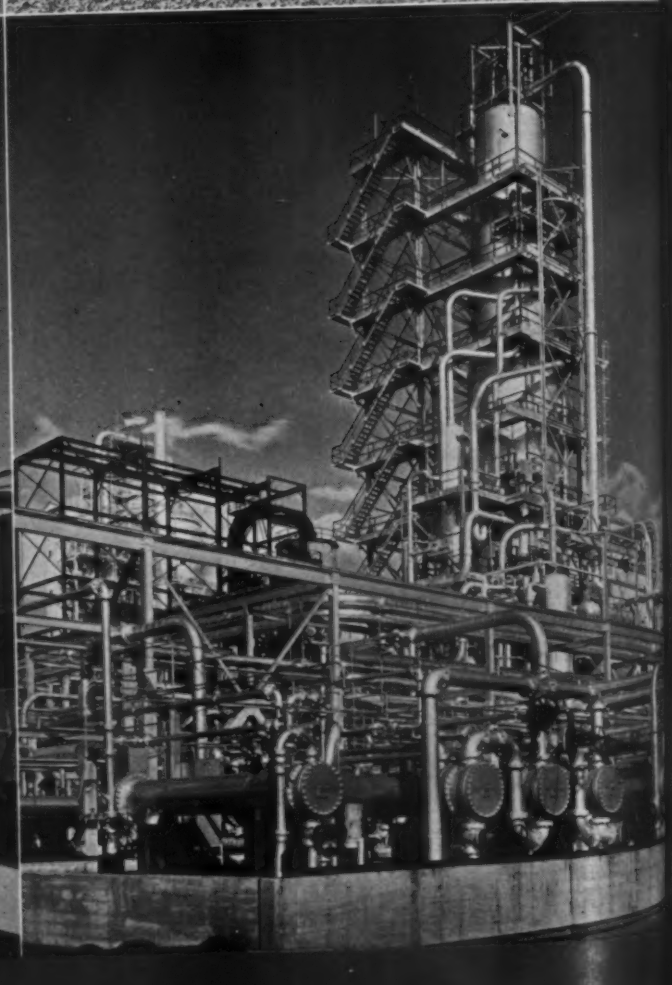

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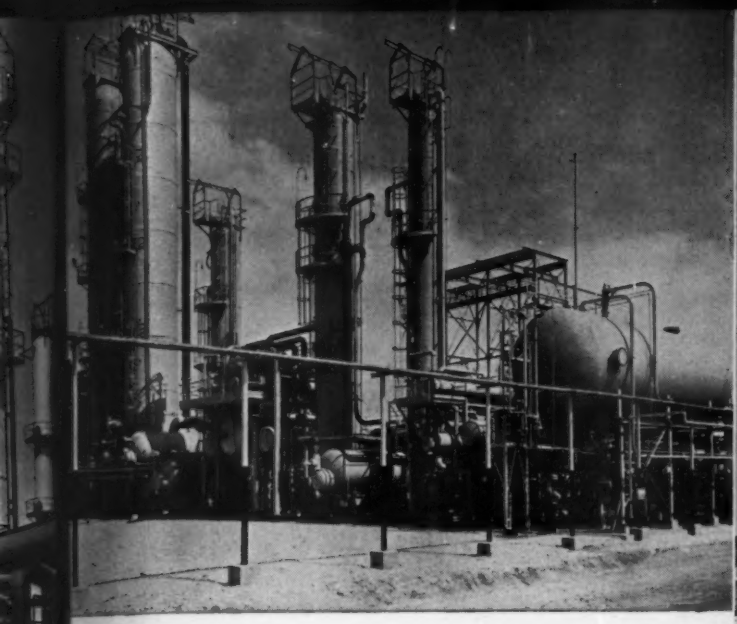

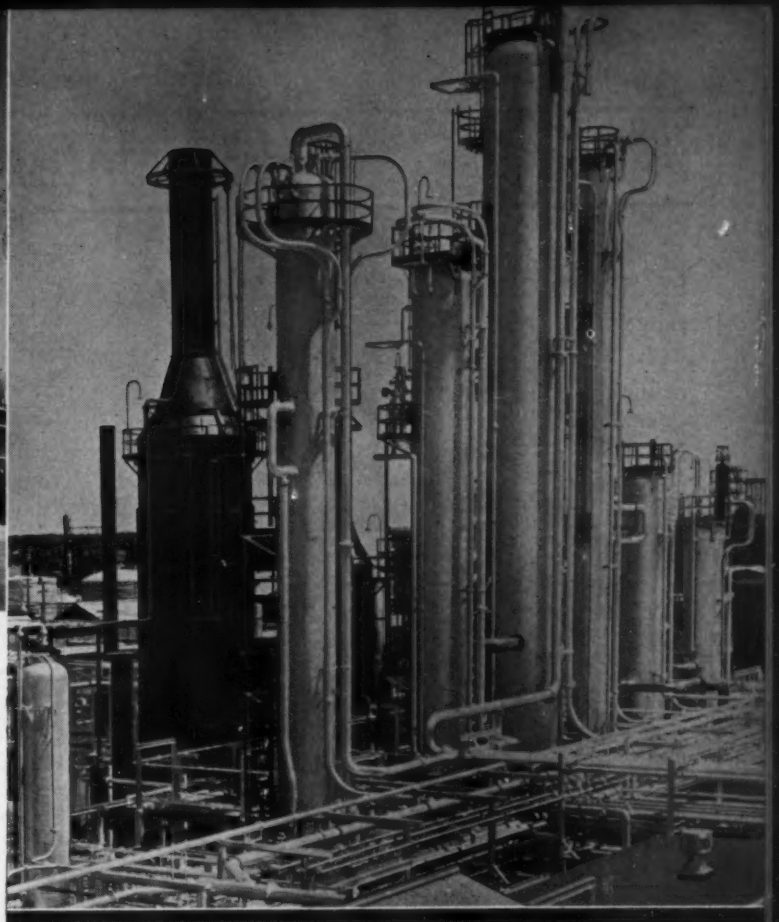

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- Lubricating Oil Refining
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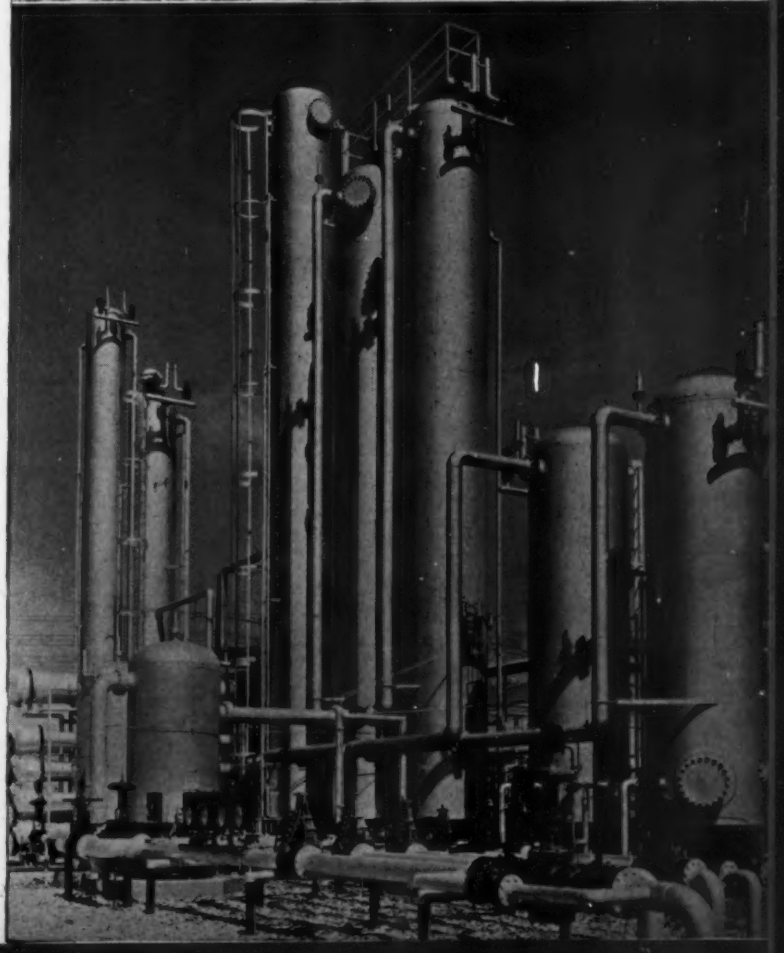

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UNDIVIDED RESPONSIBILITY IN ONE ORGANIZATION

Fluid Catalytic Cracking Process

STANDARD OIL DEVELOPMENT COMPANY

THE Fluid catalytic cracking process is a basic refinery tool of great importance in modern refining operations. This process has come into existence primarily because of its ability to produce greater yields of gasoline of a higher octane number than the thermal cracking process, which it has largely displaced. The Fluid technique is also being adapted for conversion of natural gas into liquid products, where it represents an advancement over the Fischer-Tropsch process. Technique of this process has wide potential application in other petroleum refining processes as well as in the chemical and other industries.

In the Fluid process, finely divided catalyst is suspended in gas or vapor so that it flows through the equipment with flow characteristics resembling a liquid. Thus, it circulates continuously between reaction and regeneration zones. The heat from the regenerator is transferred to the reactor, thereby replacing the need for cooling surface in the regenerator and reducing the preheat requirement for the feed.

Catalyst circulation is maintained without mechanical means by the use of standpipes. The column of catalyst in the standpipe is maintained in a fluidized state by addition of aeration gas, so controlled that the density is substantially greater than the catalyst vapor mixture in the connecting riser to the adjoining vessel. Due to the difference in densities of the catalyst-vapor mixtures in the standpipes and risers, circulation is made possible. By operating the vessels at relatively low velocity, a bed of dense solids is formed due to their tendency to settle which makes it possible to obtain the necessary catalyst holdup in vessels of moderate size.

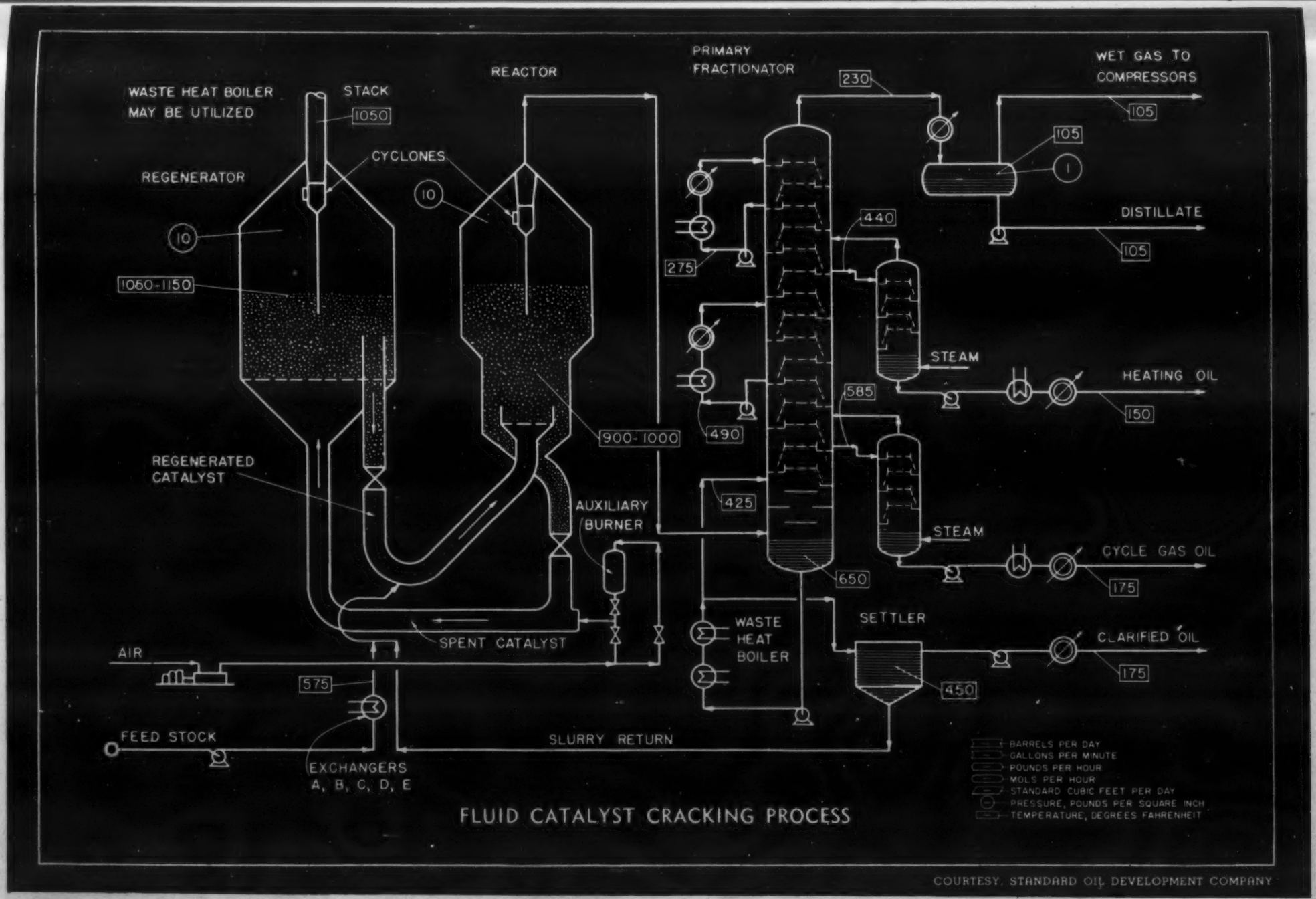
With reference to the flow diagram, oil feed is injected into the hot regenerated catalyst, which supplies the required heat. The catalyst rate is regulated to maintain required reactor temperature. This catalyst-vapor mixture flows into the reactor, where the catalyst begins to settle, forming a dense bed, which resembles a violently boiling liquid with a rather clearly defined level. Vapors leave the top of the reactor through separators which return most of the entrained catalyst to the bed. Cracked products then are processed in conventional fractionating equipment, in which any remaining catalyst is concentrated for return to the reactor.

In the reaction step, the catalyst accumulates a carbonaceous deposit in the form of coke. The spent catalyst is removed from the reactor through a stripping zone, for removal of retained hydrocarbons, and transferred in suspension with air to the regenerator, where the coke is removed by combustion. The heat released is absorbed by the circulating catalyst. Flue gases leave the regenerator through cyclones, where the catalyst content is reduced to a low value, and may be cooled in conventional waste-heat boilers.

The process data shown are for a typical gas oil operation. High yields of motor gasoline are obtained with clear octane numbers of 81-83 ASTM and 92-99 CFR. By-products, such as isobutane and light olefins for alkylation, polymerization, LPG or for synthetic-rubber ingredients are recovered.

Feed stocks in commercial operation range from naphtha cuts to heavy gas oils and deasphalted residuums. Recycling operations have been found attractive in some instances. Flexibility of the process has been demonstrated through uninterrupted runs of more than a year, and by general high-service-factor levels. High catalyst recovery has been established in commercial plants, based on a circulation rate corresponding to a loss averaging about 0.25 pounds per barrel of feed.

Fluid process units now operating or under design and construction throughout the world represent an operating capacity of approximately 1,400,000 barrels per stream day, or more than half the total catalytic cracking capacity. Capacity of the units ranges from 2,000 to 41,000 barrels per stream day.



FLUID CATALYST CRACKING PROCESS

COURTESY, STANDARD OIL DEVELOPMENT COMPANY

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Climax Engineering Co. Controls Division

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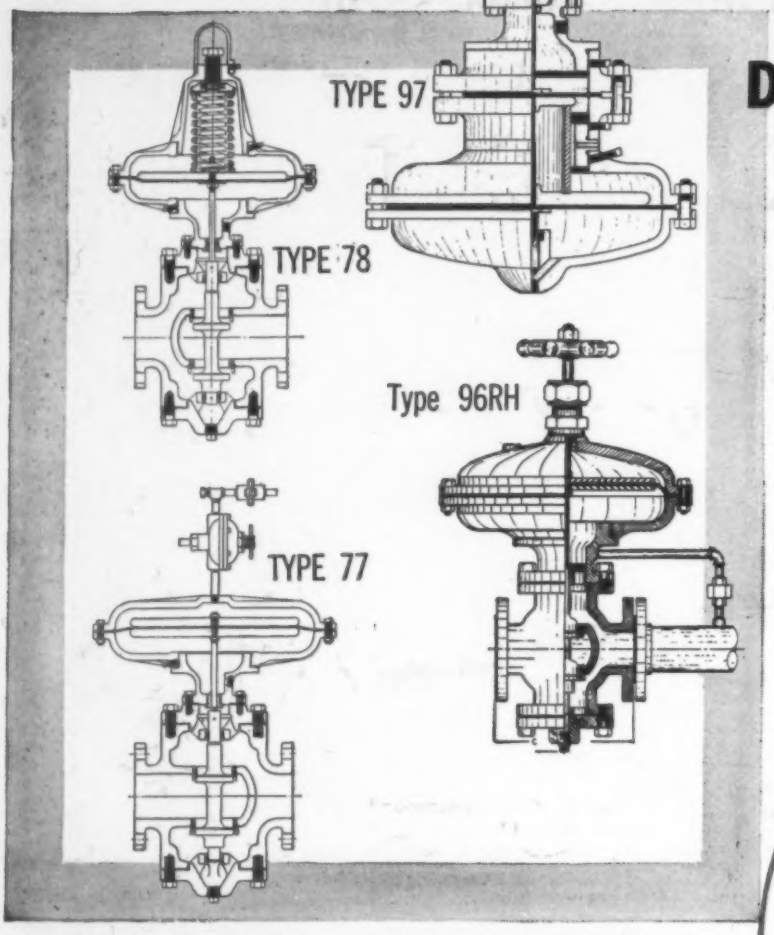

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Rugged, dependable, precise . . . they are made for superior service with liquids or gases. Ideal for back-pressure, pressure reduction, fuel lines, steam service line applications; scores of other purposes. Write for complete description, specifications.

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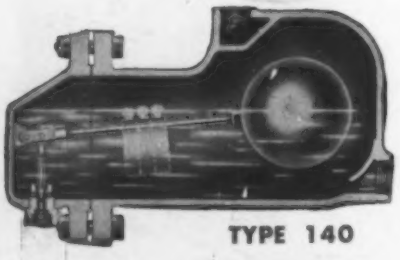
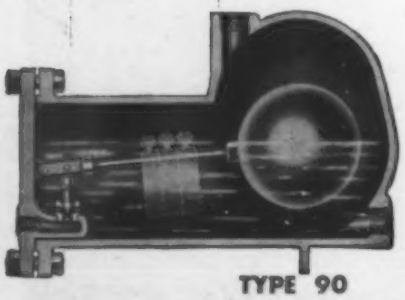


TYPE 140



TYPE 90

Achieve two-phase continuous drainage of water, condensate or oils from pressure systems. Ideal for use as steam traps on fractionators, reboilers, kettles, other equipment using steam for heat.

Type 90 is for systems operating at pressures up to 500 psi. Type 140 is for high working pressures. It can also be employed as a level control on high gravity and corrosive liquids. It, too, can be used in inverted position . . . as an air, gas or vapor eliminator from pressure systems. Write today for full information.

Patent No. 2,263,771

The Original

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For external mounting on vessels, kettles, separators, receivers and other units where you can isolate the controller without shutting down system. Float sizes from 14 to 120 inches. Cast steel housings for pressures up to 1000 psi; fabricated steel for higher pressures. Special types available for corrosive service.

Incorporates famous Climax Pilot Controller for on-off or up to 300% throttling range. Overcomes "hunting" action, assures *lined-out* control.

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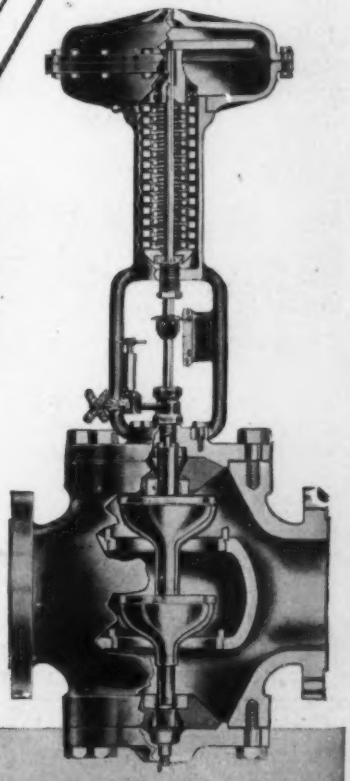
"High-lift," "characterized" plugs for wide rangeability and proportional flow increase — large diaphragm areas and high spring loadings minimize "hysteresis." All double-port and single-port reverse structures fully reversible in field.



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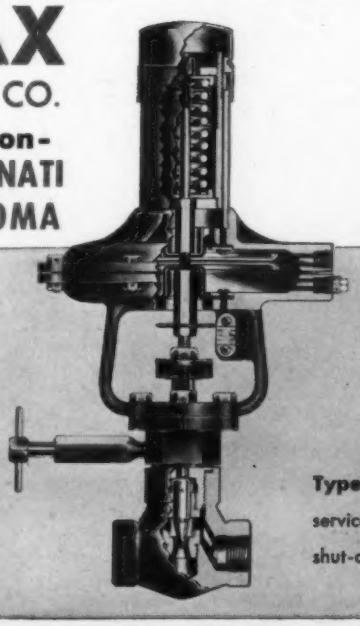
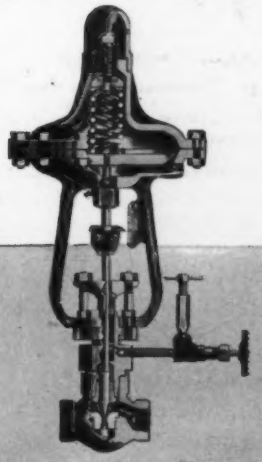


Type 1515—Especially designed for use in pilot plants where small size is a requirement. Unique, patented reversible top-works provides for easy change from normally closed to normally open and the reverse, without dismantling. Screwed bonnet construction—cooling fins for temperatures over 450°F.



Type 15DRT Designed for high pressure service. Single-port construction and positive shut-off—ideal for high pressure gas lines.

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Fluid Catalytic Cracking Process

THE M. W. KELLOGG COMPANY

THE primary purpose of the Fluid Process as with other petroleum cracking operations is to break down long-chain and high-molecular-weight hydrocarbons into molecules boiling within the gasoline range. The Fluid Process differs from other catalytic-cracking processes in that the reaction is effected by a *powdered* catalyst which is circulated essentially as a fluid with the oil to be cracked. Advantages of the system as a means of controlling and handling catalyst are:

(1) A highly turbulent bed in the reactor provides intimate contact between catalyst and vapor.

(2) The ability to control reaction conditions closely over a wide range produces optimum yields with uniform quality of products.

(3) Continuous uniform regeneration is obtained with minimum degradation of catalyst activity.

(4) The fluid catalyst may be circulated freely without regard to limitations of mechanical transfer equipment. Flows are maintained by self-perpetuating hydrostatic heads.

(5) Heat is conveyed from the regeneration system to the reaction system without the danger of thermal cracking and coking.

Hot catalyst at a temperature of approximately 1100° F. passes from the regenerator into the regenerator standpipe down which it flows to the oil-injection point where it flash vaporizes the fresh-oil charge which has been pre-heated to balance the total heat requirements of the reaction.

The mixture of oil vapors and catalyst is forced up the reactor-feed line by the pressure head maintained in the regenerator plus the force of gravity exerted on the dense catalyst in the catalyst standpipe, the pump serving merely to bring the oil to the oil-injection point.

The hot mixture enters the reactor where the catalyst settles to a definite level and forms a "bed," depth of which regulates the time of reaction, and can be varied at will. The bed is maintained in a fluid, turbulent condition by the entering feed vapors. Continually passing upward, these vapors effect intimate contact of oil with catalyst and produce a uniform temperature (900° - 950° F.) throughout the bed. As cracking progresses, coke forms on the catalyst.

The spent catalyst, laden with coke, is continually and automatically withdrawn through a slide valve to the spent-catalyst stripper where the adsorbed and entrained vapors are dissociated from the catalyst and carried by a counter-current stripping medium back into the top of the reactor vapor space. Spent catalyst separates from the oil vapors by gravity, drops into the spent-catalyst standpipe and through an automatically controlled slide valve to the air-injection point where its density is reduced by a current of air which "fluffs up" the catalyst so that it flows as a fluid up the regenerator return line.

The oxygen in the air admitted for "fluffing up" purposes continuously regenerates the catalyst by burning off coke. This action brings the catalyst to a temperature of about 1100° F. and makes it an effective heat-transfer medium for vaporizing the fresh-oil feed and supplying the heat necessary for the cracking reaction. The reactivated hot catalyst is retained in the regenerator and is constantly withdrawn into the standpipe for re-use as needed.

The combustion gases pass through cyclone separators in the top of the regenerator where they follow a high-velocity spiral path which throws the heavier catalyst particles to the outside of the separator by centrifugal force. The recovered catalyst is returned to the bed while the flue gases enter the regenerator exit line and pass through the flue-gas cooler where their contained heat is used to generate high-pressure steam by heat exchange with water.

The cooled flue gases then pass to the electrical precipitator. Flue gases are vented to the atmosphere. Recovered catalyst passes from the precipitator into the recovered-catalyst line and is blown back to the regenerator.

The combined hydrocarbon gases from the spent-catalyst stripper and the reactor join in the vapor space above the catalyst bed and enter cyclone separators where catalyst is recovered and returned to the main bed. The cracked gases pass through the reactor overhead line to the fractionator scrubbing section where the small quantity of residual catalyst fines is diverted to the bottom of the fractionator and passes with the heaviest oil to the slurry settler, an integral part of the fractionator in which the catalyst settles out by gravity. The clarified decanted oil is withdrawn from the top of the settler for use as fuel oil or for further processing.

The thickened catalyst slurry passes into the catalyst-return line and is returned to the reactor. This completes the circuit of catalyst in the system.

The vapor stream entering the fractionator passes upward from the scrubbing section through the bubble plates of the fractionator and usually is separated into wet gases, unstabilized gasoline, light gasoil, and heavy gasoil.

The heavy gasoil, or the entire fractionator side stream, can be diverted for a recycling operation to the recycle-oil line and so returned to the reactor for further cracking with the fresh feed.

The table shows results obtained from single-pass operation at 50 percent conversion level on five representative feed stocks.

Comparison of Single Pass Yields at 50 Percent Conversion Gasoils from Representative Crude Sources. Synthetic Catalyst. 900° F.

	West Texas	Coastal	Mid-Cont.	East Venez.	Arab.
Product Inspections—					
Propylene in Propanes—Vol. Percent	73	70	75	70	75
Butane-Butene Fraction:					
Butenes—Vol. Percent	58	55	58	48	60
i-Butane in Sats.—Vol. Percent	80	80	83	83	83
Debutanized Motor Gasoline:					
Reid Vapor Pressure—lb.	6.1	6.2	6.3	6.5	7.0
Gravity, °API	55.1	55.0	56.9	56.0	57.0
ASTM Percent (D plus L) at 158 °F.	27	27	28	27	28
212	50	50	51	49	54
284	68	70	73	70	70
356	88	90	91	90	90
Octane Numbers—					
ASTM—Clear	81.1	82.0	82.0	80.5	83.0
CFRR—Clear	94.6	95.0	95.1	94.2	96.0
Total Cycle Oil:					
Gravity, °API	24.5	23.0	26.8	25.0	26.0
ASTM 50 Percent Pt. °F.	600	610	604	580	585

Licenses for this process are obtainable from The M. W. Kellogg Company.

KELLOGG FLUID CATALYTIC CRACKING PROCESS
REGENERATOR
REACTOR
FRACTIONATOR
GAS TO RECOVERY UNIT
GAS SEPARATOR
UNSTABILIZED GASOLINE
LIGHT GAS OIL
HEAVY GAS OIL
DECANTED OIL
WATER
REFLUX
CYCLONE SEPARATORS
VAPORS
SPENT CATALYST STRIPPER
STRIPPING STEAM
CATALYST
SLURRY SETTLER
CATALYST RETURN
RECYCLING IF DESIRED
AIR
STEAM
WATER
FLUE GAS COOLER
ELECTRICAL PRECIPITATOR
FLUE GAS
RECOVERED CATALYST
PREHEAT FURNACE
CATALYST
OIL FEED
COURTESY, THE M. W. KELLOGG COMPANY

One of the several jobs engineered and built by M. W. Kellogg under a master contract . . . complete refinery in the Rocky Mountain area.

The "Master" Contract – GOOD OR BAD?

EVEN before the last war, a definite trend seems to have been established which has become accelerated in recent years—in favor of the so-called "master" contract.

Under this type of contract *all* engineering and construction work to be undertaken at any particular refinery site is placed with *one* engineering contractor—instead of being distributed among several.

From our experience—in the building of complete domestic refineries at Denver, Lake Charles, Salt Lake City and currently at East Chicago, Indiana, plus complete foreign installations made in Palestine and currently contracted for in Brazil, Canada, Argentina and the British Isles—there *are* definite and direct advantages in this method of operation which are probably the reason for the increasing number of "master" contracts being let.

For instance, the "master" type contract offers to the refiner—big or small,

1. *Concentration of overall process responsibility*
   ... with consequent increase of efficiency of final plant and more profitable employment of refiner's own staff
2. *Integrated process planning*
   ... providing actual initial capital savings and continued improved operating efficiency
3. *Coordinated mechanical engineering*
   ... eliminating costly hours of coordinating time and job site alterations
4. *Uniformity of engineering specifications*
   ... providing economy in maintenance and spare parts inventory
5. *Centralized purchasing of materials and equipment*
   ... permits allotment of scarce materials in accord with desired completion schedule
6. *Simplified field administration and accounting*
   ... removing the necessity of duplicate facilities by several contractors in the field and the concurrent extra work at home offices
7. *Elimination of competition for local labor*
   ... expediting construction and insuring minimum labor costs
8. *Uniform labor policy at job site*
   ... with resultant improvement of man-hour efficiency
9. *Maximum economy in use of all labor at job site*
   ... through the flexible interchange of crafts from one unit to another
10. *Saving in temporary facilities at job site*
    ... by dispensing with duplicate utility lines and housing for tools, supplies and labor
11. *Reduced tool expense*
    ... through full time use of all tools on various units

by
Warren L. Smith
Vice President of The M. W. Kellogg Co.

Detailed discussion of each of these advantages accruing to the refiner who selects the "master" contract operation, by specialists in various phases of refining engineering and construction, are available in printed form on request.

Our company—M. W. Kellogg—is completely staffed and equipped for refinery and chemical plant design and construction, from the step of laboratory research through the initial operation of completed plants. It periodically issues a publication called the KELLOGGRAM, which describes many of these services. If you are not already receiving it, we will be glad to add your name to the list if you will just drop us a line at 225 Broadway, New York 7, N. Y.

M KELLOGG W
A SUBSIDIARY OF PULLMAN, INC.

NEW YORK JERSEY CITY LOS ANGELES TULSA HOUSTON TORONTO LONDON PARIS

Fluid Catalytic Cracking Process

UNIVERSAL OIL PRODUCTS COMPANY

THE fluid catalytic cracking process is adaptable to the needs of large and small refiners. Units designed by Universal Oil Products Company, in capacities ranging from 2600 to 18,000 barrels a day, have been in successful operation for years. UOP improved fluid catalytic cracking units can be built and operated at costs comparing favorably with the cost of thermal cracking units of comparable capacity.

Development and commercialization of the fluid catalytic cracking process marks an important forward step in petroleum refining. The process produces gasoline of a different character from that produced by thermal cracking.

The process is flexible both as to charging stock and products. It can successfully crack kerosine, light or heavy gasoil and some types of reduced crude. It can be operated to produce either aviation gasoline, high grade motor gasoline or raw materials for the production of chemicals and synthetic rubber.

Principal modifications in the improved unit as compared to the units of earlier design are:

1. In the small size units the reactor and regenerator have been combined in a single tower which is erected as a self-supported column. The heavy structure required in the earlier units has been eliminated.

2. The large regenerator riser has been eliminated, and the regenerator standpipe has been considerably shortened. There has been a general reduction and simplification of small piping.

3. Efficiency of the reactor and regenerator has been improved by the use of additional distribution grids.

4. The air supplied for combustion passes through the regenerator countercurrent to the catalyst. This permits multistage stripping and more complete utilization of oxygen.

The oil feed is charged as a liquid at the base of the reactor riser, while hot catalyst from the regenerator enters the riser at the point of oil injection, supplies the process heat requirements and is transported upward by the vaporized oil into the reactor. Spent catalyst is continuously withdrawn from the upper part of the reactor dense phase to hold a level in the reactor, and passes downward through the stripping section into the regenerator. Three zones are available for stripping in this design. The stripping medium is injected at the bottom of both the internal and the external strippers. Additional stripping of the spent catalyst is obtained in the upper section of the regenerator where the bulk of the flue gas from the combustion of coke in the lower sections contacts the incoming catalyst. The catalyst passes downward through the regenerator countercurrent to the air supplied at the bottom for regeneration.

Regenerated catalyst flows from the lower section of the regenerator to the reactor riser at a rate controlled to maintain a constant reactor temperature. The regenerator flue gas is vented to the atmosphere through an internally mounted cyclone separator which recovers entrained catalyst. The hydrocarbon products from the reactor pass to the fractionator likewise through an internally-mounted cyclone separator which recovers entrained catalyst and returns it to the catalyst bed.

By building the regenerator below the reactor it is now possible to pass the air for regeneration countercurrent to the flow of catalyst. With the countercurrent flow and staging of the regenerator with grid plates the oxygen content of the outlet gas can be reduced to ½ to 1 percent while in previous units two to three times as much unused oxygen normally escapes from the regenerator. This feature of the new design was not possible in the older units.

The latest type unit can utilize either synthetic or natural catalyst. In its most simple design synthetic catalyst which has a microspherical form is used. Because of the spherical shape of the particle and its greater strength as compared with the ground powdered catalysts used during the war, this catalyst has very low erosion and attrition rates. In addition, the M.S. catalyst is much more uniform in particle size than the ground catalyst. Because of the reduced formation of fines by attrition, plants using M.S. catalyst can dispense with secondary catalyst recovery on the gas from the regenerator while still maintaining a low catalyst loss.

Economic consideration of market requirements and operating costs will govern the choice of catalyst. The silica-alumina catalyst gives the highest octane number but a lower gasoline yield, while somewhat more gasoline having a lower octane number is produced from the natural catalyst.

The table below shows the yields obtained in pilot plant once-through and recycle cracking of 31° API Mid-Continent gasoil over silica-alumina catalyst.

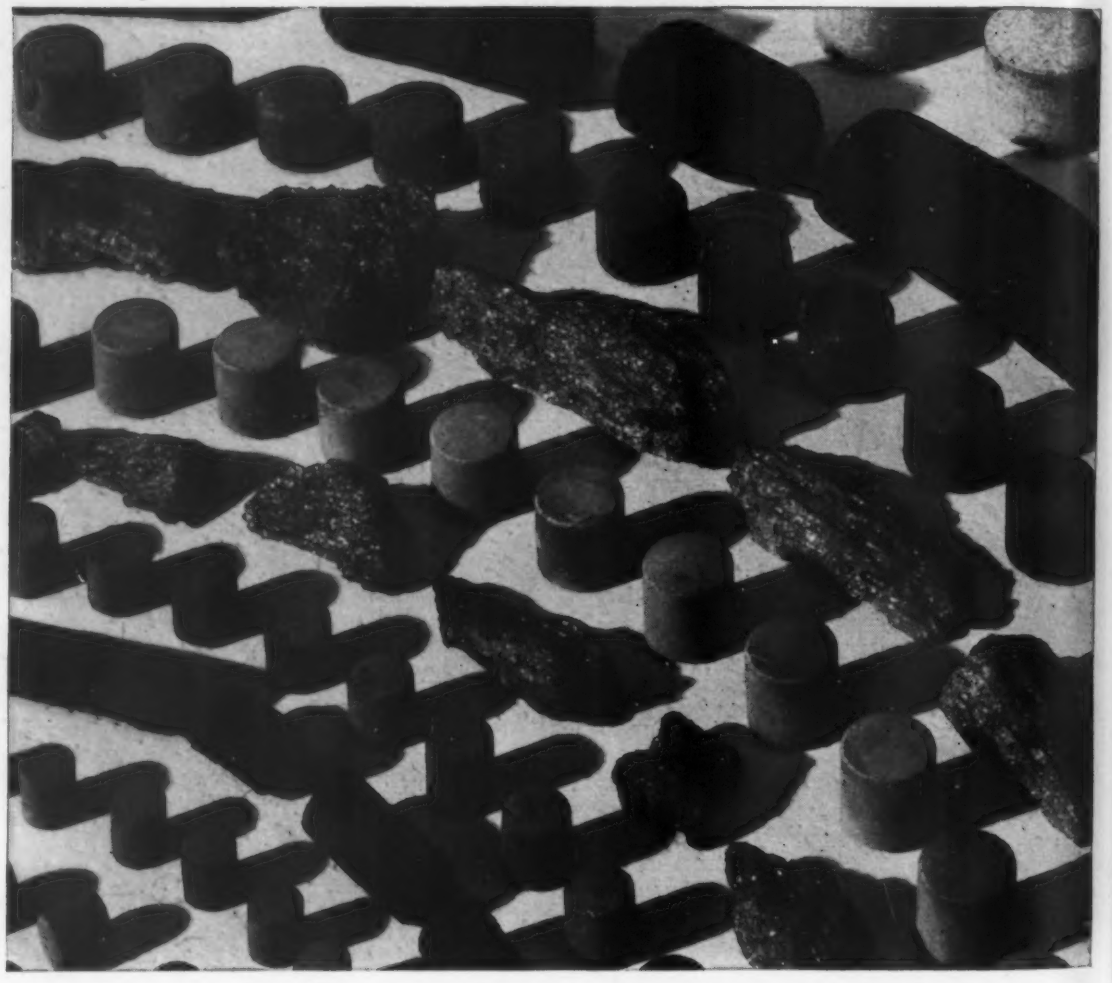
Fluid Catalytic Cracking 31° API Mid-Continent Gasoil (Silica-Alumina Catalyst) Pilot Plant Results

OPERATION	Once-Through	Recycle
Conversion, percent	60	75
Yields, Volume, percent:		
Debutanized gasoline	45.0	53.0
Cycle oil	40.0	25.0
Butylenes	5.8	7.1
Butanes	8.7	9.4
Propylene	5.2	6.4
Propane	2.4	3.3
C_2 and lighter gas, weight percent	2.5	3.1
10 lb. RVP catalytic gasoline	47.4	55.8
10 lb RVP polymer gasoline	8.0	9.9
Total 10 lb. RVP gasoline	55.4	65.7
Octane Numbers of 10 lb. Catalytic Gasoline:		
Motor method clear (F-2)	81.8	81.8
+ 3 cc TEL/Gal.	86.8	87.4
Research method clear (F-1)	93.7	94.2
+ 3 cc TEL/Gal.	98.7	99.1

FLUID CATALYTIC CRACKING PROCESS
COURTESY, UNIVERSAL OIL PRODUCTS COMPANY
SILENCER
REACTOR
CYCLONE
STRIPPER SECTION
STRIPPER MEDIUM
STRIPPER MEDIUM
CYCLONE
REGENERATOR
AIR HEATER
AIR COMPRESSOR
CATALYST ADDITION
STEAM
WATER
FLUE GAS STEAM GENERATOR
CATALYST WITHDRAWAL
COOLER
CLARIFIED SLURRY
STEAM
RAW OIL CHARGE
SLURRY SETTLER
STEAM GENERATOR
WATER
MAIN COLUMN
STEAM
COOLER
SIDE CUT STRIPPER
STRIPPER
SIDE CUT
SIDE CUT
GASOLINE
CONDENSER
DISTILLATE RECEIVER AND GAS SURGE DRUM
GAS COMPRESSOR
GAS

USE HARSHAW

PREFORMED CATALYSTS



Recent developments in the field of organic chemistry have resulted in many new processes which have made necessary a wide variety of catalysts. Harshaw pioneered in the development of preformed catalysts. Not only do we have wide experience in the development and use of tabletted and extruded catalysts, but in addition, we have a large capacity for producing them. If you have a catalyst problem, a discussion with us may be helpful.

THE HARSHAW CHEMICAL CO. • CLEVELAND, OHIO

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AND CATALYTIC CHEMICALS

for

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CYCLIZATION
OXIDATION
DEHYDROGENATION
HYDROGENATION
DEHYDRATION
DESULPHURIZATION
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HARSHAW

TYPICAL CATALYSTS

Aluminum Chloride Anhydrous
Boron Trifluoride
Boron Fluoride Addition Compounds
Hydrogen Chloride Anhydrous
Hydrofluoric Acid Anhydrous

Activated Alumina	Tungsten Alumina	Magnesia
Chrome Alumina	Cobalt	Nickel
Molybdena Alumina	Copper	Vanadium

We offer our manufacturing facilities and the collaboration of our research staff for developing catalysts fitted to your various processes.

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Fixed-Bed Catalytic Cracking Process

HOUDRY PROCESS CORPORATION

IN THE field of catalytic cracking the Houdry process employing fixed catalyst beds is the pioneer. The first commercial-scale plant was put in operation in 1936 and since then many additional units have been installed in both the United States and abroad.

Two fundamental needs of catalytic cracking are to furnish heat during the endothermic cracking period and to remove heat during the exothermic regeneration, or the period during which the carbonaceous catalyst deposit formed from the cracking reaction is burned. The fixed-bed type reactor or case has been designed to fulfill these requirements.

Essentially, the reactors are tubular heat-exchange vessels. The catalyst bed lies in the shell side through which flow the oil or air vapors. On the tube side a molten mixture of alkali metal salts is circulated. Inasmuch as the net effect of catalytic cracking and carbon burn-off is exothermic, the circulated salt is used as a cooling medium. The high heat capacities of the steel reactor structure and of the catalyst bed have been major factors in design of equipment.

Full utilization is made of the large quantities of heat evolved during operation. The bulk of the heat is removed from the reactors by the flowing molten salt, of which a part is circulated through a by-pass to the salt-steam generator and superheater and another portion through a salt-air heat exchanger used for preheating the regeneration air to the cases. The relatively cool salt returning from the boiler and air heater is fed again into the main body of salt, and by this means the total salt to the reactors is maintained constant at any desired temperature. The heat carried from the reactors by the regeneration gases is utilized in providing power to the gas turbine used to drive the compressor supplying regeneration air.

Major improvements in the Houdry fixed-bed process have been affected during the past three years by redesigning the reactor tubes and internal arrangements. The heat-exchange conditions have been improved and mechanical and process advantages have been obtained. Since the new design is a substantial simplification of the original, lower operating costs result by virtue of reduced maintenance and downtime for catalyst change and turn-arounds. All vapor distributing elements are now easily removable from the reactors and may, therefore, be removed and thoroughly cleaned on every catalyst change, assuring good distribution of the vapors and maximum yields of the desired cracked products. Because of the improved mechanical distribution thus attained and the higher on-stream efficiencies, greater production of high-octane gasoline results.

Four 12-case Houdry units have been converted to the new design to date and engineering is presently underway for revamping six additional units to the new design. Experience in the operation of the modified reactors during the past three years indicates that not only are operating costs reduced and gasoline output increased, but improved gasoline quality has been achieved because of a greater supply of heat available within the catalyst mixture for the cracking reaction.

The flow diagram shows the three essential parts of the Houdry system—(1) a furnace for vaporization of the charge to the reactors, (2) the catalytic or case section, and (3) the fractionating equipment for splitting the synthetic crude into useful products. Only three catalyst cases are shown, illustrating the least number generally employed in Houdry cracking plants. Larger plants employ multiples of three or four cases with the reactors manifolded in pairs.

An electrical timer transposes the cyclic operation of each reactor to a continuous operation of the overall unit. The cycle is, briefly, that one reactor or pair is on-stream while a second is being regenerated and the third is on oil purge, air purge, or repressuring operation and undergoing the necessary valve changes. The purges are affected by steam-jet ejectors and, in addition, during the oil evacuation, sweeping steam is admitted to the reactor. Motor-operated valves actuated by impulses from the cycle timer assure continuous oil and air flow through the system, and appropriate electrical interlocks preclude any mixing of the streams.

By adjustment of the case operating variables such as temperature, pressure, space rate, etc., the relative yields of gasoline, fuel gas and catalytic gasoil may be varied over a wide range. Typical operating conditions, yields, and inspections when cracking gasoil for motor gasoline production are given in the accompanying table.

Typical Houdry Fixed-Bed Operation Cracking Gasoil from Mixed Base Type Crude

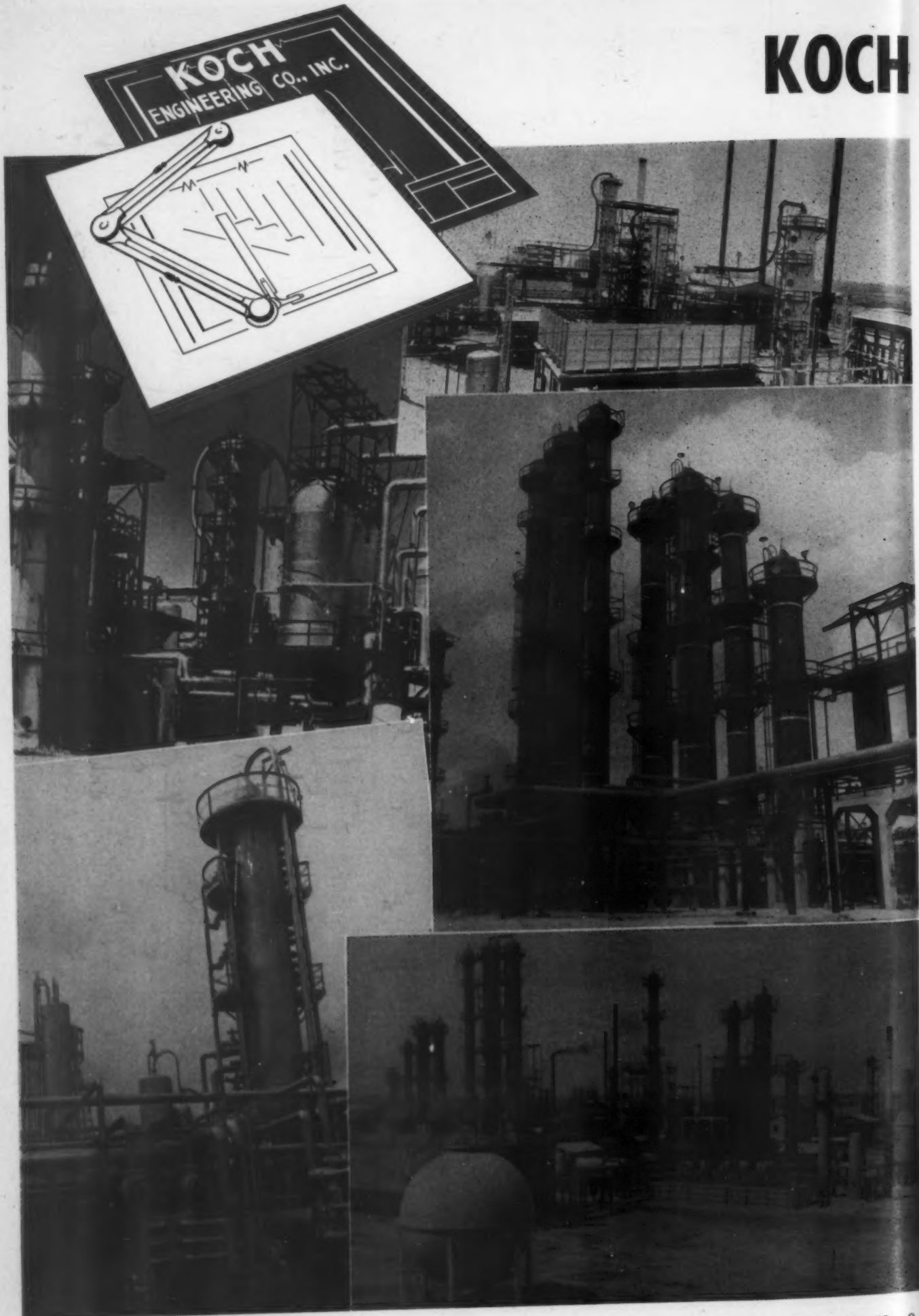
Type Process—Once-Through Cracking for Motor Gasoline Production
Type Charge Stock—Gasoil from Mixed-Base Type Crude
Type Catalyst—Synthetic

Operating Conditions:

Space Velocity, V/Hr/V	1.0
Temperature of Salt to Reactors, °F	810
On-Stream Reactor Pressure, psig	15
Process Steam	None
Time on Stream, Minutes	10
Time of Complete Cycle, Minutes	30

	Vol. Percent Fresh Charge to Reactor	Gravity °API	5 Percent	10 Percent	50 Percent	90 Percent	95 Percent	R.V.P. (psi)	C.F.R.M. Clear Octane	Saybolt Color
Charge:										
Virgin Gasoil	100.0	35.2	490	506	560	672	710	....	....	....
Products:										
Motor Gasoline	46.3	57.9	115	127	232	368	....	9.9	78.9	30
Catalytic No. 2 Fuel Oil	39.8	32.5	470	480	520	590	610	....	....	....
Catalytic Heavy Gasoil	5.0	18.0	588	600	650	710	....	....	....	....
Excess C_4 Fraction	9.9	.....	....	....	....	.....	....	....	....	....
Dry Gas, C_3 & Lighter, Weight Percent	4.8	....	....	....	....	....	....	....	....	....

WET GAS TO STABILIZER
MOTOR GASOLINE TO STABILIZER
LIGHT CATALYTIC GAS OIL
HEAVY CATALYTIC GAS OIL
CHARGE STOCK
SYNTHETIC TOWER
HEAT EXCHANGER
AIR HEATER
CATALYST CASES
TURBO-COMPRESSOR
CHARGE HEATER
FRESH AIR
SPENT AIR TO STACK
HOUDRY CATALYTIC CRACKING PROCESS
SIMPLIFIED FLOW DIAGRAM
COURTESY, HOUDRY PROCESS CORPORATION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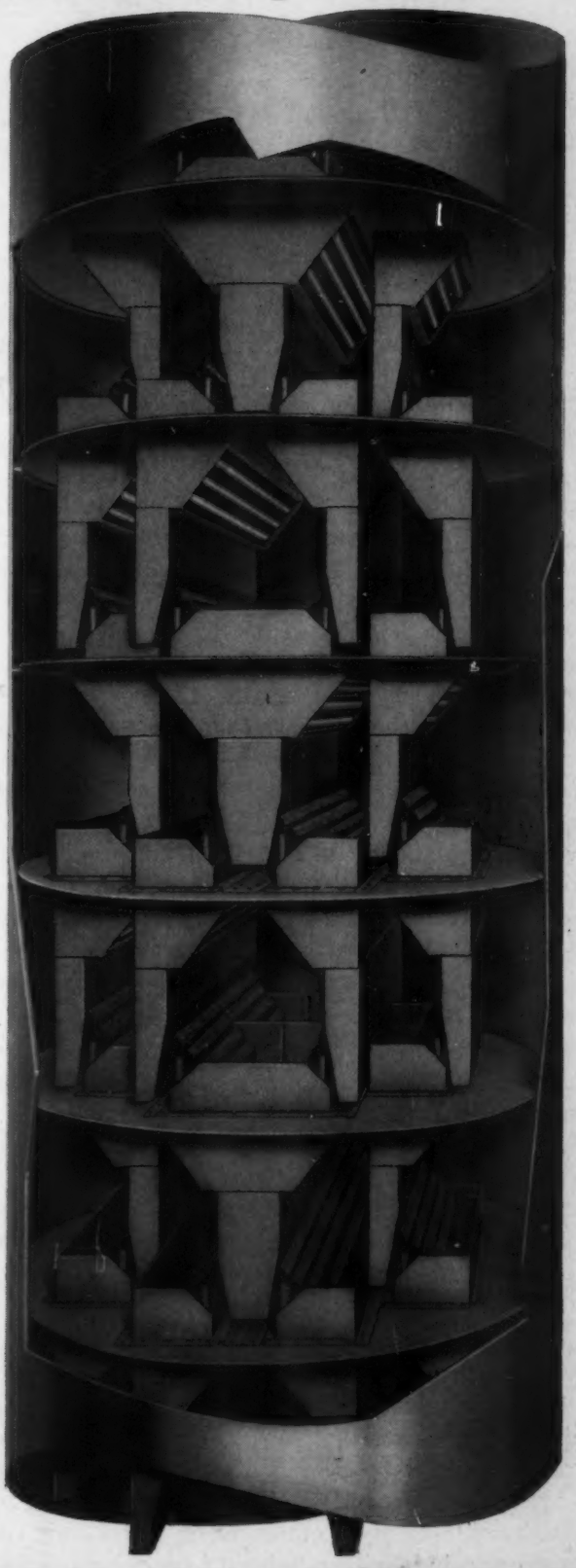

Illustrated at the right are Koch Kaskade fractionating trays installed in a 700 pound absorber.

Large vapor and liquid areas of the Koch Kaskade trays afford 35 to 100 percent greater capacity with less pressure drop than bubble trays. Better contacting permits using only 80 percent as many Koch Kaskade trays as bubble trays for the same separation.

The Koch Kaskade tray sections are bolted to a liquid distributing plate. These sections are easily inserted or removed through a manway.

Prompt deliveries can be made in any size from 2' 4" diameter to any larger diameter desired, of carbon steel, stainless steels, Monel, copper or aluminum construction.

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Cycloversion Process

PHILLIPS PETROLEUM COMPANY

CYCLOVERSION or catalytic cracking developed by Phillips Petroleum Company, Bartlesville, Oklahoma, is available for license to the petroleum refining industry through the Perco Division.

This process is schematically similar to the Perco Catalytic Desulfurization and Catalytic Reforming Process elsewhere described in this Process Handbook, and as a consequence Cycloversion may be used to accomplish desulfurization, catalytic reforming and catalytic cracking.

The operation of the Cycloversion Process when catalytically cracking gasoil, is illustrated in the accompanying diagram. The fresh or recycle feed is charged to a heater where it is preheated to about 980-1000° F. A small amount of steam diluent is added to the feed in the convection bank of the furnace to reduce carbon deposition on the catalyst, improve the octane rating and assist in vaporizing the charge stock. The superheated oil-steam mixture passes thence to the catalyst chamber at about 85 psig where catalytic cracking is obtained in process cycles of periods of 2-10 hours. Waste-heat exchangers are provided at the catalyst-chamber outlet to recover a large amount of heat in steam production. From the waste-heat exchangers the product passes to the heat exchange and condensing equipment and onto the oil-water-vapor separator where the condensed water is withdrawn, the vapors go to the recovery plant and the liquid hydrocarbon condensate to fractionation.

Simultaneous with the processing of the hydrocarbon charge in the on-stream reactor, the alternate catalyst chamber is undergoing regeneration. Steam from the waste-heat exchangers is superheated to about 850° F. admixed with air and charged to the chamber on regeneration. The regeneration gases are passed to waste-heat exchangers and finally vented to the atmosphere through a suitable vent stack.

The Cycloversion Process embodies a fixed-bed principle of catalytic cracking in which no appreciable loss of catalyst through the formation of fines by abrasion occurs. Equipment for the circulation, separation, and recovery of catalyst from the process or reactivation gases is not required. The rugged natural bauxite catalyst employed is capable of regeneration an indefinite number of times and maintains 85-95 percent of the cracking efficiency of the new catalyst. Other features of the Cycloversion Process are its low-cost operation together with high yields of superior-quality motor fuels. In addition, the flexibility of the process lends it to catalytic reforming of motor-fuel-boiling-range naphthas as well as desulfurizing both cracked and straight-run products.

Typical Cycloversion Process results are presented in the accompanying tables.

Low Pressure Cycloversion of Texas Panhandle Gas Oil*

	Undesulfurized	Desulfurized
Feed Stock:		
Gravity, API°	36.9	....
10 Percent	552	....
90 Percent	697	....
Aniline Point, °F	185	....
Catalytic Gasoline, C_5-400° F.	**Test 1**	**Test 2**
Gravity, API°	57.2	56.1
Reid Vapor Pressure	6.4	6.4
ASTM Octane Number	81.5	81.2
+ 3 cc TEL	85.3	84.9
Research Octane Number	94.3	94.4
+ 3 cc TEL	98.0	97.8
Yield, Percent of Charge:		
Catalytic Gasoline, C_5-400° F. Vol. Percent	28.1	35.0
10 Lb. RVP Gasoline, Vol. Percent†	38.7	47.3
Propylene, Vol., Percent	5.0	6.0
Butylenes, Vol. Percent	7.3	8.0
Catalytic Recycle, 400° F. +, Vol. Percent	57.8	40.5
Carbon, Wt. Percent	1.2	2.7
Conversion, Single Pass, Vol. Percent	42.2	59.5

* 10 Psig operating pressure.

† Butanized and includes potential catalytic poly gasoline.

High Pressure Cycloversion of Texas Panhandle Gas Oil*

Feed Stock:		
Gravity, API°	37.4	....
10 Percent	501	....
90 Percent	709	....
Aniline Point, °F	180	....
Catalytic Gasoline, C_5-400° F.	**Test 1**	**Test 2**
Gravity, API°	56.3	55.4
Reid Vapor Pressure	6.4	4.8
ASTM Octane Number	76.7	77.7
+ 3 cc TEL	82.4	83.4
Research Octane Number	87.1	89.2
+ 3 cc TEL	93.4	94.8
Yields, Percent of Charge:		
Catalytic Gasoline, C_5-400° F. Vol. Percent	33.3	34.0
10 Lb. RVP Gasoline, Vol. Percent†	40.9	44.2
Propylene, Vol. Percent	3.7	4.9
Butylenes, Vol. Percent	4.3	5.4
Catalytic Recycle 400° F. +, Vol. Percent	53.6	45.0
Carbon, Weight Percent	1.5	2.6
Conversion, Single Pass, Vol. Percent	46.4	55.0

* 85 Psig operating pressure.

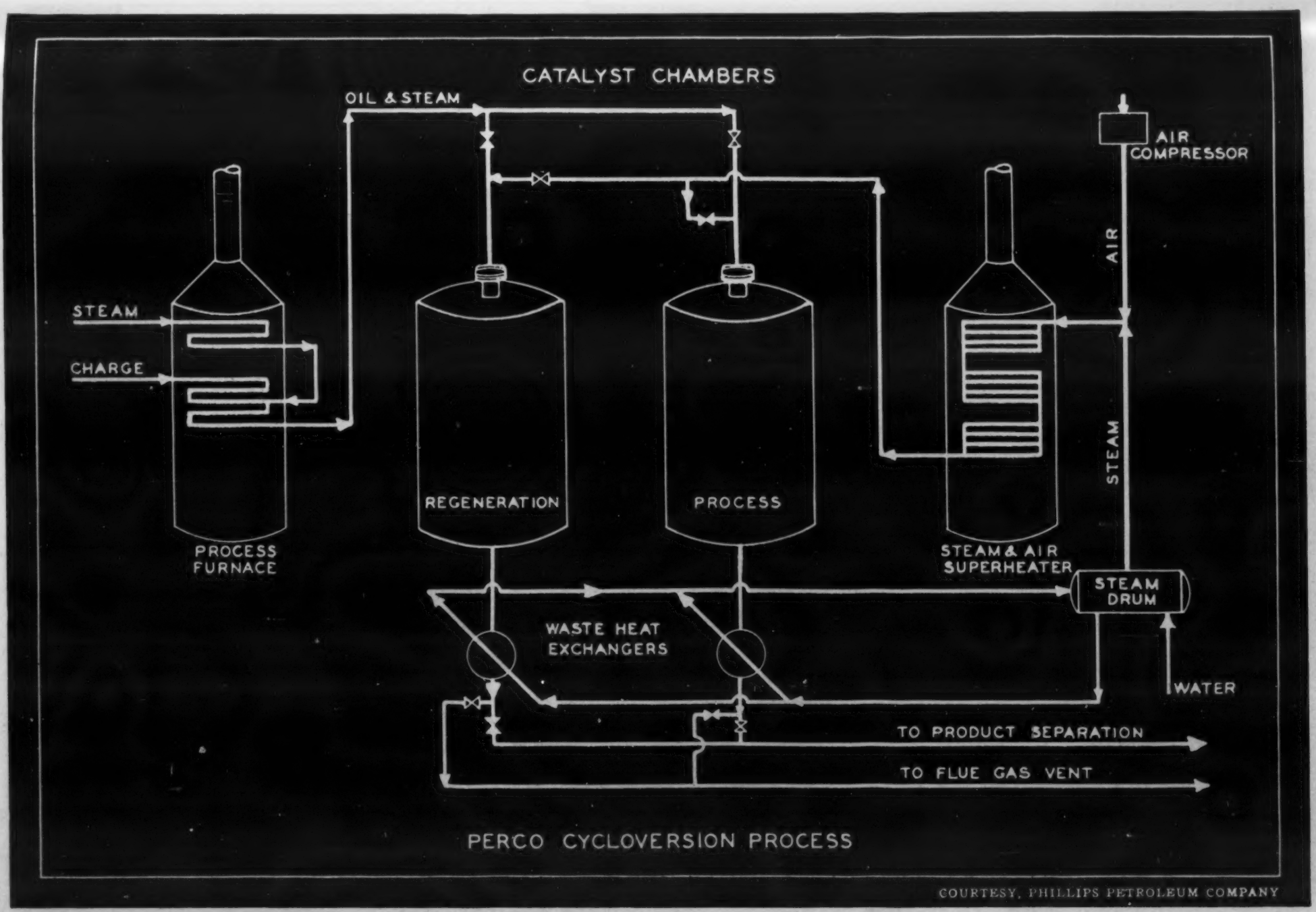
† Butanized and includes potential catalytic poly gasoline.

Low Pressure Cycloversion of Mid-Continent (Western Kansas) Gas Oil*

Feed Stock:		
Gravity, API°	32.2	....
10 Percent (Vacuum)	604	....
90 Percent (Vacuum)	752	....
Aniline Point, °F	181	....
Catalytic Gasoline, C_5-400° F.	**Test 1**	**Test 2**
Gravity, API°	55.1	54.9
Reid Vapor Pressure	6.0	6.6
ASTM Octane Number	80.4	81.1
+ 3 cc TEL	84.5	84.7
Research Octane Number	94.3	94.5
+ 3 cc TEL	97.8	98.0
Yields, Percent of Charge:		
Catalytic Gasoline, C_5-400° F., Vol. Percent	28.6	33.2
10 Lb. RVP., Gasoline, Vol. Percent†	37.4	42.9
Propylene, Vol. Percent	4.1	4.2
Butylenes, Vol. Percent	4.9	5.6
Catalytic Recycle 400° F. + Vol. Percent	59.1	50.7
Carbon, Weight Percent	1.5	2.3
Conversion, Single Pass, Vol. Percent	40.9	49.3

* 10 Psig operating pressure.

† Butanized and includes potential catalytic poly gasoline.



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For Flow Chart and description see Pages 72 and 73 this issue.

3 words that ADDED 5 TIMES THE LIFE to Heat Exchanger Tubes

Photo by Vachon. Courtesy Standard Oil Co., (N.J.)

Every seven months, sulfuric acid ate away the heat exchanger tubes in one refinery.

Frequent replacement of equipment proved expensive. So refinery engineers began a series of tests with other corrosion-resistant metals.

Tests prove helpful

On the basis of test results, the engineers made this recommendation: *"Change to Monel*."*

Now—with Monel tubes—the refinery gets 43 months of life *in the same service where other tubes averaged only seven.* (See Co-Polymer Plant history in panel.)

Performances like this aren't unusual. Nor are they confined to sulfuric acid applications.

Many refinery uses

Monel withstands a wide range of refinery corrosives. For example:

- *hydrogen sulfide*
- *hydrochloric acid*
- *sodium hydroxide*
- *ammonia*
- *impure or brackish waters*
- *oil, wax and solvent mix*

Monel also resists hydrofluoric acid in HF alkylation; it serves dependably under high-temperature, high-pressure conditions in feed water heaters; withstands attack in caustic soda reboilers and in condensers for chlorinated solvents and furfural.

Long life, low cost

Wherever corrosive conditions are severe, consider using Monel. It may help you increase equipment life, speed up production, and cut maintenance and replacement expense.

"Change to Monel" PAYS DIVIDENDS

Co-Polymer Plant

Monel tube bundle installed in reactor feed precooler. (121 tubes, ¾" OD x 180" x 14 BWG.)

Product Handled —Butane and 70% sulfuric acid outside tubes; butane inside.

Temperatures — Entrance to tubes: 115°F.
Exit from tubes: 145°F.
Entrance to outside: 160°F.
Exit from outside: 150°F.

Service Life —Previous material lasted 7 months; Monel tubes 43 months.

Ketone Dewaxing Unit

Monel tube bundle located in charge cooler. (754 tubes, ¾" OD x 144" x 16 BWG.)

Product Handled — Oil, wax and solvent mix outside tubes; water inside.

Temperatures — Entrance to tubes: 90°F.
Exit from tubes: 120°F.
Entrance to outside: 180°F.
Exit from outside: 110°F.

Service Life —Previous material lasted 31½ months. Monel tubes in service 67 months to date. Still in good condition when inspected. No plugged tubes.

Caustic Plant

Monel tube bundle located in second effect, caustic evaporator. (712 tubes, 2" OD x 48½" x 16 BWG.)

Product Handled — Steam from first effect outside tubes; caustic soda and sodium chloride inside.

Temperatures — Entrance to tubes: 172°F
Exit from tubes: 172°F.
Entrance to outside: 240°F.
Exit from outside: 240°F.

Service Life —All but 15 of the original 721 Monel tubes installed in 1928 are still in service. (The 15 were replaced because of failure of lap-weld seam.) Previously-used material had an average life of 10 months.

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Catalytic Reforming Process

PHILLIPS PETROLEUM COMPANY

THE Perco Catalytic Reforming Process developed by Phillips Petroleum Company, Bartlesville, Oklahoma, and licensed to the industry through its Perco Division, provides an economical means of both desulfurizing and catalytically reforming cracked and straight-run naphtha.

The normal benefits derived from Perco Catalytic Desulfurization elsewhere described in this Process Handbook, are obtained together with an appreciable increase in clear octane rating due to some additional mild cracking or reforming as evidenced by the production of fixed gases in the reactor effluent stream.

Normal process cycles are of the order of 24 hours or more with only a small deposition of carbon or other carbonaceous matter on the catalyst as compared to conventional catalytic cracking. This low carbon deposition on the catalyst permits an operation on relatively long process cycles and low-cost regeneration of the catalyst.

The process operates at low pressure, usually of the order of 50-100 psig. As shown in the accompanying flow sheets, the feed is vaporized, usually by heat exchange with the reactor effluent, and then superheated to a temperature of about 1000° F. The superheated vapors are passed through the reactor, thence to exchange with the feed and finally to product cooler and accumulator.

The regeneration of the catalyst is the same procedure as that described for Perco Catalytic Desulfurization and comprises burning off the carbonaceous deposits with a superheated air-steam mixture.

Typical Perco Catalytic Reforming results are given in the accompanying tables. Note the appreciable increase in "jump" octanes or the spread between ASTM and Research Octane, a feature of extreme interest to refiners having only thermal-type equipment. It is of further interest to note that the process is not only applicable to stocks having moderately high total sulfur but also to those stocks having relatively low sulfur.

Fixed gas losses are relatively low and yields are high, particularly in view of the C_4 and C_3 components recoverable as poly-feed.

The process may be operated either as a catalytic reforming unit or as a catalytic desulfurization unit.

ILLINOIS STRAIGHT-RUN GASOLINE
64.4° API Gravity; 90-360° F. Boiling Range

	Raw Feed	Cat. Reformed[1]
ASTM Octane No.	54.1	61.3
+1 cc TEL	64.7	73.1
+3 cc TEL	73.0	81.8
Research Octane No.	52.7	63.8
+1 cc TEL	64.2	75.0
+3 cc TEL	74.2	83.4
RVP at 100° F.	10.1	10.0
Total Sulfur, Percent	0.022	0.008

PENNSYLVANIA STRAIGHT-RUN GASOLINE
62.2° API Gravity; 100-400° F. Boiling Range

	Raw Feed	Cat. Reformed[1]
ASTM Octane No.	48.8	61.5
+1 cc TEL	57.3	71.0
+3 cc TEL	67.1	79.6
Research Octane No.	49.6	65.8
+1 cc TEL	57.9	74.4
+3 cc TEL	68.7	83.5
RVP at 100° F.	8.5	10.2
Total Sulfur, Percent	0.055	0.005

MID-CONTINENT (KANSAS) CRACKED-STRAIGHT-RUN GASOLINE
61.7° API Gravity; 121-403° F. Boiling Range

	Raw Feed	Cat. Reformed[1]
ASTM Octane No.	59.6	70.8
+1 cc TEL	68.4	78.8
+3 cc TEL	75.6	83.6
Research Octane No.	63.2	75.3
+1 cc TEL	71.4	83.6
+3 cc TEL	81.2	88.8
RVP at 100° F.	4.4	10.6
Total Sulfur, Percent	0.039	0.0 1

PANHANDLE CRACKED-STRAIGHT-RUN GASOLINE
59.2° API Gravity; 117-390° F. Boiling Range

	Raw Feed	Cat. Reformed[1]
ASTM Octane No.	57.2	62.7
+1 cc TEL	63.8	71.4
+3 cc TEL	71.6	77.7
Research Octane No.	59.5	67.1
+1 cc TEL	66.9	7.7
+3 cc TEL	74.8	83.3
RVP at 100° F.	4.8	4.1
Total Sulfur, Percent	0.082	0.032

EAST TEXAS DISTILLATE (2)
67.1° API Gravity; 100-410° F. Boiling Range

	Raw Feed	Cat. Reformed[1]
ASTM Octane No.	47.8	70.7
+1 cc TEL	59.9	78.4
+3 cc TEL	73.0	83.3
Research Octane No.	47.0	75.7
+1 cc TEL	57.1	85.1
+3 cc TEL	72.8	90.8
RVP at 100° F.	6.0	8.0

(1) Catalytically reformed products are Cr and heavier gasolines and do not conform potential poly gasoline from polymerization of Cr and Cr olefines produced.
(2) Sulfur content of feed and product not reported.

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HYDROCARBON FLOW
REGENERATION GAS FLOW
STEAM
AIR
STEAM SUPERHEATER
TO VENT STACK
CATALYST CHAMBERS
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GAS TO VAPOR RECOVERY
PRODUCT TO SEPARATION
PROCESS HEATER
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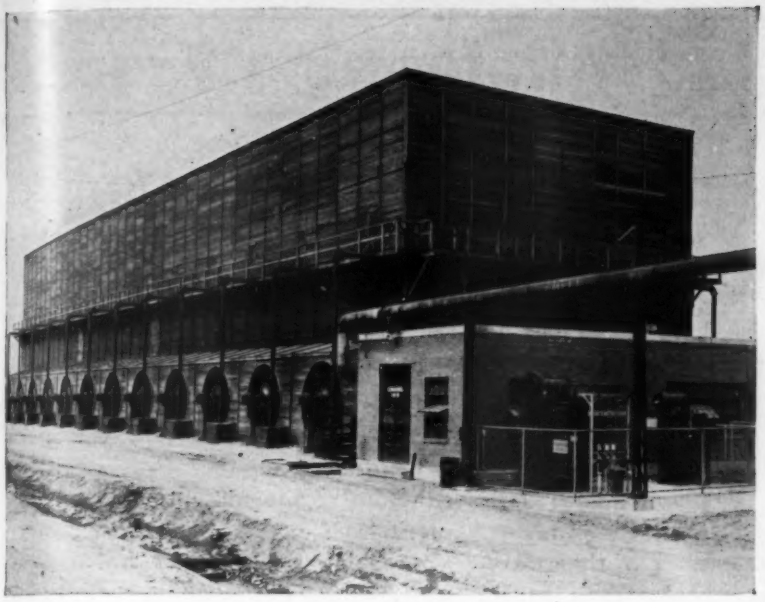

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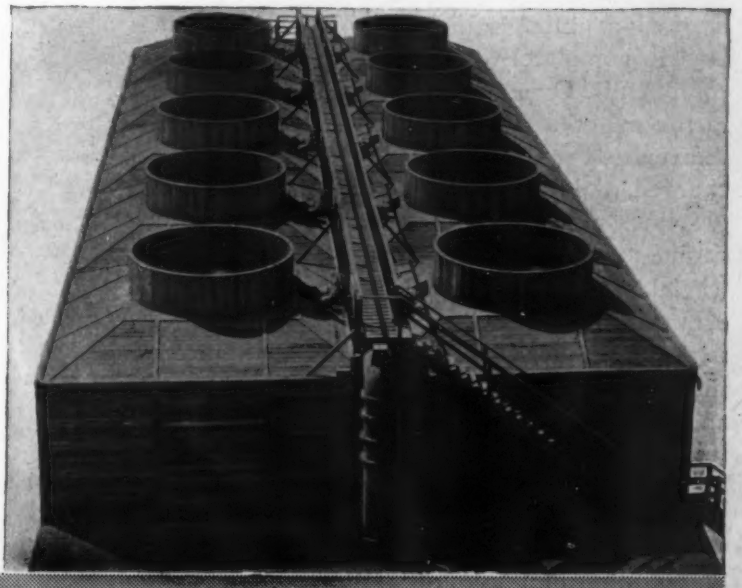

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Hydroforming Process

THE M. W. KELLOGG COMPANY

HYDROFORMING is primarily a catalytic process for converting a low-octane naphtha into a high-octane gasoline suitable for fuel in high-compression automobile engines. This conversion or reforming is accomplished largely by means of dehydrogenation and cyclization reactions. These reactions are promoted by molybdena-on-alumina catalyst at temperatures of 900-1000° F. and pressures of 150-300 psig with a high partial pressure of hydrogen in the reaction zone. The hydrogen partial pressure is maintained by recycling hydrogen-rich gas to the reaction zone along with the vaporized fresh feed. The process usually produces a large amount of net hydrogen, which makes possible the hydrogen-rich recycle.

Naphthenes (cycloparaffins) in the feed stock are dehydrogenated, and some molecules are slightly rearranged to form aromatics. Considerable cyclization of straight-chain hydrocarbons also takes place. If any olefins are present, as is the case when same cracked naphtha is included in the feed, these are almost completely converted to paraffins or ring compounds. Over 90 percent of the sulfur contained in the naphtha feed is removed, giving a low-sulfur reformate even when a sulfur-bearing naphtha is charged. Additional characteristics of hydroformer gasolines are low Reid vapor pressures; high clear octane numbers; high lead susceptibilities.

Yields of 78-80 volume percent of 80 CFR-M octane, 100 percent C_4 recovery gasoline are obtained, with the balance of the material going approximately three fourths to C_4-free gas and one fourth to polymer and carbon. When operating to lower octane levels it is possible by blending in extraneous butanes to the 10-pound Reid vapor pressure level, to obtain a yield based on feed of 100 percent of 69-octane gasoline which can be brought up to 80 CFR-M octane by the addition of 1.5 cc. of TEL.

Hydroforming is carried out in a unit usually having four or more reactors filled with catalyst, half of which are always in series on reaction, and the other half in different stages of regeneration. The reactors are switched from reaction through regeneration and back to reaction in cycles of 8 to 16 hours duration by an electric cycle-controller. Regeneration is necessary because a small deposit of carbon and a partial reduction of the metal in the catalyst combine over a period of hours to reduce the activity of the catalyst. The carbon is burned and the catalyst reoxidized by combustion in a stream of air greatly diluted by cooled and recirculated flue gas.

The reaction products are recovered and fractionated into: (1) a hydrogen-containing gas substantially free of hydrocarbons heavier than propane, (2) a highly aromatic "polymer" boiling above 400° F., and (3) a depropanized, 400° F. end-point gasoline.

The following table shows the inspections of a typical naphtha feedstock together with the inspections of the gasoline produced from it.

	FEED West Texas and Mid-Continent Mixed Naphtha		Hydroformer Gasoline
Gravity, °API	50.5		50.5
ASTM, Distillation, °F:			
IBP	174		104
10 Percent	236		156
50 Percent	311		258
90 Percent	384		338
EP	442		399
Octane Number, CFR-M clear	39.7		80.3
Aniline Point, °F	124		57
Bromine Number	2.2		3.2
Lamp Sulfur, Weight Percent	0.220		0.015
Yields, Vol. Percent of Feed (Output Basis):			
C_4-Free Gasoline		73.6	
Butanes		10.6	
Polymer		3.2	
Yields, Wt. Percent of Feed Output Basis:			
C_4-Free Gasoline		76.0	
Butanes		7.8	
Polymer		3.9	
C_4-Free Gas		10.7	
Carbon		1.6	
Total		100.0	

KELLOGG HYDROFORMING PROCESS
COURTESY, THE M. W. KELLOGG COMPANY
STEAM
WASTE FLUE GAS
AIR
COMPRESSOR
FIRST PASS REACTOR
COMBINATION FURNACE
DOWN FLOW BURNING
FLUE GAS RECIRCULATOR
SECOND PASS REACTOR
UP FLOW BURNING
WASTE HEAT BOILERS
FEED WATER
H2 ENRICHMENT SECTION of FRACTIONATING SYSTEM
RECYCLE GAS COMPRESSOR
HEAT EXCHANGE WITH FRACTIONATION SYSTEM
GAS SEPARATOR
TO MAKE GAS ABSORBER
HYDROFORMATE TO FRACTIONATING SYSTEM
NAPHTHA FEED
PC
FC
TC

FOR Good SERVICE
Look for
the Green
Triangle

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The Continental organization was founded on the precepts that its success and growth depended upon giving the oil and gas industries a representative line of *quality* equipment . . . and *good* service. Continental's steady, yet unhurried, expansion has proved this policy's soundness.

Today "The Green Triangle" is a familiar—and looked for—sign throughout the oil country.

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Hydrogenation Process

STANDARD OIL DEVELOPMENT COMPANY

HIGH-pressure hydrogenation for oil processing and the methane-steam "tube" method for manufacture of hydrogen used in hydrogenation were developed in the United States by Standard Oil Development Company. Plants utilizing this process have been operated at Baton Rouge, Louisiana; Bayway, New Jersey; Port Arthur, Texas; Richmond, California, and Whiting, Indiana, as well as at various locations abroad. In addition, "tube" manufacture of hydrogen has been adapted to a number of relatively small installations, while over 100 million cubic feet per day of "tube" hydrogen capacity was installed in government plants for ammonia production during the war.

High-pressure hydrogenation has been used in manufacture of aviation base stock and motor gasoline from gas oils, high-grade diesel fuels from low-quality gas oils, water-white paraffinic kerosenes from inferior-quality distillates, and high-viscosity-index lubricating oils from poor-quality lubricating distillates. The process also lends itself to preparation of high solvency naphthas, high-octane-number aviation safety fuels, etc. During the war large quantities of isooctane aviation blending agent were produced using high-pressure hydrogenation; an unusual adaptation was its use to make aromatic amine anti-knock addition agent for aviation gasoline from the corresponding nitro compound. High-pressure hydrogenation plants can be used for desulfurization of high-sulfur fuels, and for saturation of catalytically cracked cycle stocks to permit their recycle cracking to ultimate high production of catalytic gasoline.

In the hydrogenation process the charging stock, together with sufficient hydrogen, is pumped into a reaction vessel containing catalyst. At the desired temperature and at about 3000 psi, the reaction takes place. The "tube" method for hydrogen manufacture and the hydrogenation step proper are outlined in the accompany flow sheet.

Hydrogen may be produced by any of the usual processes. Since the hydrogen does not need to be sulfur-free, its production from coal or coke by the water-gas process is facilitated; it may also be produced by liquefaction and distillation of coke oven gases, or electrolytically, when the economics permit. But for the oil industry with its generous supply of refinery and natural gas, hydrogen is advantageously made by the treatment of hydrocarbons with steam in accordance with the following reactions:

$$CH_4 + H_2O \rightarrow CO + 3H_2$$
$$CO + H_2O \rightarrow CO_2 + H_2$$

The first step is carried out in a reforming furnace, using catalyst in tubes through which reaction heat is supplied. The second, "conversion," step also uses catalyst, but does not require a supply of heat. After cooling, the carbon dioxide is scrubbed out, leaving hydrogen of about 95 percent purity.

The oil and previously compressed hydrogen are mixed and delivered through heat exchangers to a coil furnace and thence to a reaction vessel containing catalyst. The degree of hydrogenation is controlled carefully by alteration of operating conditions. From the reaction chamber the products and gases pass through the heat exchangers and coolers to a high-pressure separator where the liquid product is separated from the unconsumed hydrogen and other gases. The liquid product is reduced to atmospheric pressure and sent to storage.

Gas formed in the process ordinarily is removed in the liquid product owing to its solubility at the high pressure. Such gases as are not removed in this manner may, if necessary, be taken out by scrubbing the recycle hydrogen with oil under pressure. The unconsumed hydrogen separated under high pressure from the liquid product is recycled by compressing back to full operating pressure in a booster compressor, which operates through a pressure interval equal to the pressure drop in the system. Fresh hydrogen is added to the recycle gas to make up for that consumed in the process.

No coke is formed in high-pressure hydrogenation, and the catalysts employed are unusually rugged, so that the process is continuous. Runs of more than a years' duration have been made on full-scale units, and total catalyst life of several years has been obtained. Hydrogen manufacture catalysts are also rugged and regularly maintain satisfactory activity for a number of years, so that only occasional small make-up is required. As a result, both hydrogenation and hydrogen manufacture sections show favorable service factors, and maintenance and turnaround costs are low.

The process data shown on the accompanying flow plan covers a typical operation for converting isooctenes to isooctanes, for the production of aviation blending stock.

HYDROGENATION PROCESS
COURTESY, STANDARD OIL DEVELOPMENT COMPANY
HYDROGEN PRODUCTION
HYDROGENATION
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
GAS
PRODUCT TO DISTILLATION OR STORAGE
KNOCKOUT BOX
PRODUCT SEPARATOR
SEPARATOR
BOOSTER
COOLER
CONVERTER
STEAM
NATURAL GAS OR LOW SULPHUR REFINERY GAS
REFORMING FURNACE
SEWER
WATER
H_2 HOLDER
H.P. SCRUBBING
PRODUCT COOLER
EXCHANGER
SCRUBBER
H.P. FEED PUMP
COMPRESSOR
OVEN
FIRED COIL
FEED TANK
FEED PUMP
REACTIVATOR
CO_2
SOLVENT CO_2 SCRUBBING
SOLVENT PUMP

610½ DAYS ON STREAM "FLUID CAT CRACKING" RECORD

provides a lesson in Filtrol® catalyst supremacy!

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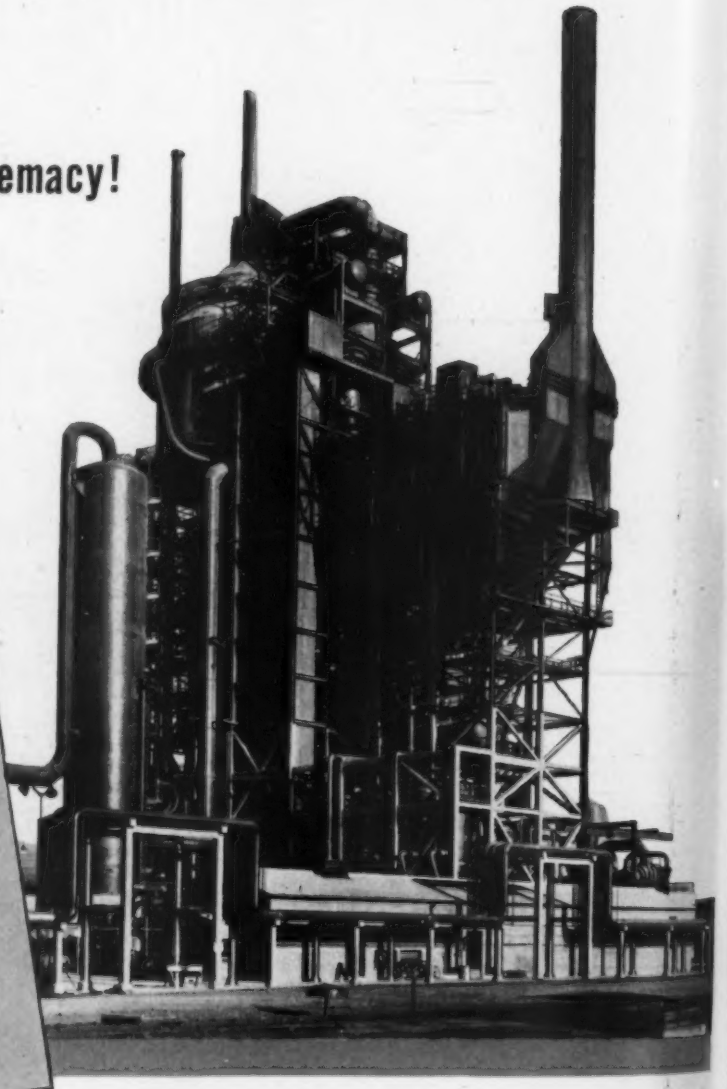

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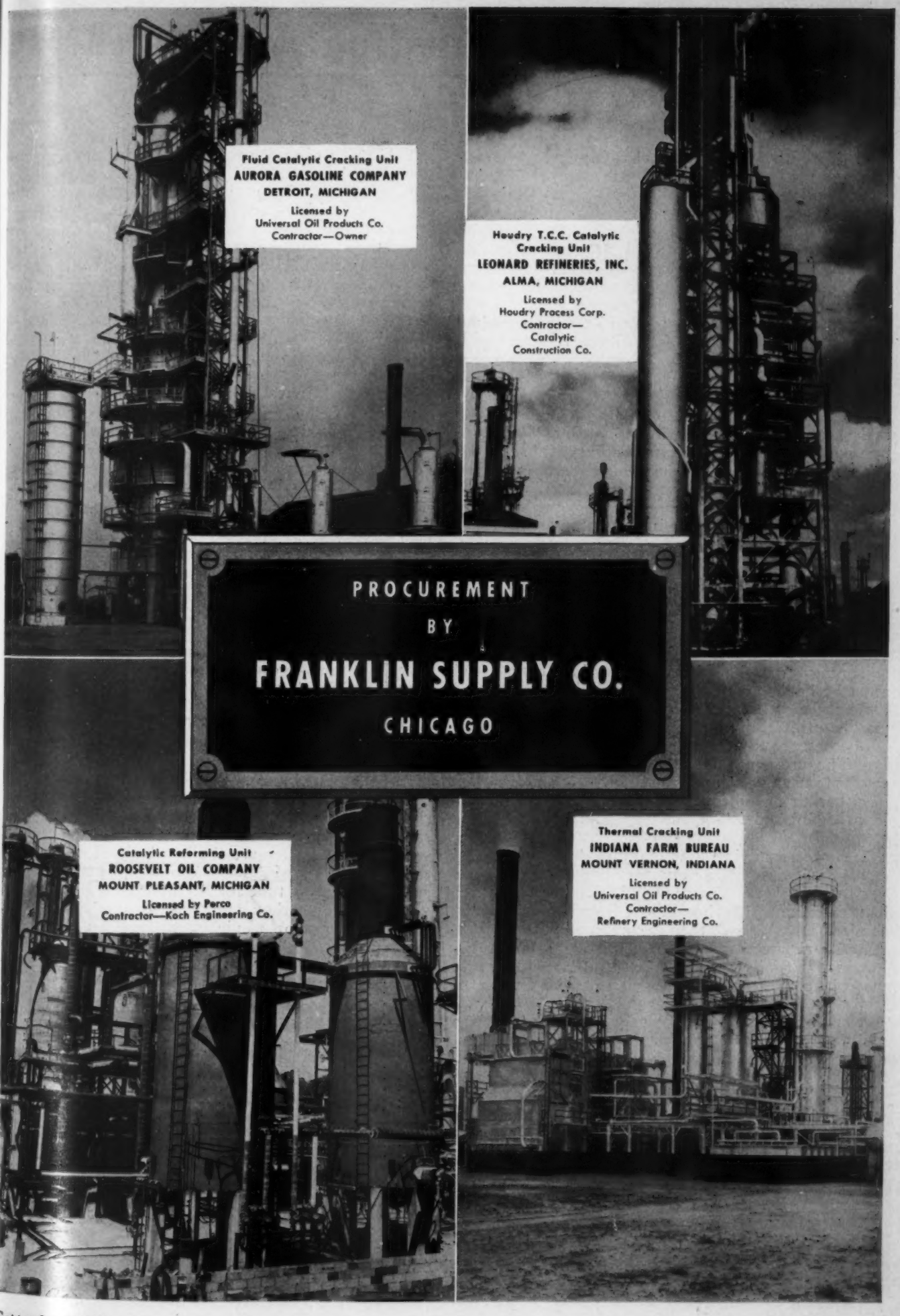

Fluid Catalytic Cracking Unit
AURORA GASOLINE COMPANY
DETROIT, MICHIGAN
Licensed by
Universal Oil Products Co.
Contractor—Owner
Houdry T.C.C. Catalytic
Cracking Unit
LEONARD REFINERIES, INC.
ALMA, MICHIGAN
Licensed by
Houdry Process Corp.
Contractor—
Catalytic
Construction Co.
PROCUREMENT
BY
FRANKLIN SUPPLY CO.
CHICAGO
Catalytic Reforming Unit
ROOSEVELT OIL COMPANY
MOUNT PLEASANT, MICHIGAN
Licensed by Porco
Contractor—Koch Engineering Co.
Thermal Cracking Unit
INDIANA FARM BUREAU
MOUNT VERNON, INDIANA
Licensed by
Universal Oil Products Co.
Contractor—
Refinery Engineering Co.

Catalytic Hydrogenation Process

SHELL DEVELOPMENT COMPANY

A SIMPLE and economical catalytic hydrogenation process has been developed by the Shell Companies and used successfully for several years for hydrogenation of octenes produced by polymerization of butylenes. The process operates at very mild conditions of temperature and pressure, effecting substantially complete hydrogenation and producing fuels low in acid-heat number suitable for manufacture of high-octane aviation gasoline.

Commercial application to date has been limited to direct hydrogenation of an octene mixture prepared by the butylene polymerization process described in this volume under the title, "Hot-Acid Polymerization." The product of these two processes is a saturated octane of 98 to 100 octane number. The hydrogenation process is by no means limited to treatment of the octene mixture described here.

The mild conditions under which this process operates are made possible by the use of a nickel catalyst supported on porcelain. The unformed catalyst is a hydrated nickel nitrate impregnated in porcelain. The nitrate is decomposed to nickel oxide, in situ, and then reduced to metallic nickel by a stream of hydrogen. After formation the catalyst is maintained in an atmosphere of hydrogen when not in use. Oxygen and carbon monoxide cause temporary loss of catalyst activity which can be restored by regeneration in place. Sulfur compounds cause permanent poisoning; thus, the sulfur content of the octene feed must be held to a minimum by proper treatment of the butylenes before polymerization. Spent catalyst may be utilized in a guard chamber for pretreatment of the octenes to prevent premature catalyst poisoning. The presence of water is not injurious to the catalyst; usually steam is used to purge the system of either air or hydrogen. A catalyst life of 1000 gallons of finished iso-octane per pound of nickel is common.

The accompanying flow diagram depicts a typical catalytic hydrogenation unit embodying four converters with two guard chambers. Subsequent to heat exchange with the effluent from the converters, the feed is vaporized, then is combined with the prereated hydrogen stream and enters the guard tubes in which partially spent catalyst is utilized to remove possible converter catalyst poisons. The pretreated feed then flows to the converters. The catalyst in the converters effects essentially complete hydrogenation and the resultant octane fraction, which is predominantly iso-octane, exchanges heat with the feed and flows to storage.

When the activity of the catalyst decreases to a certain point, it is reactivated in place. Although initial activity is not restored, two to four reactivations lengthen catalyst life by 150 percent. Activity declines slowly after each reactivation; however, the operating variables are adjusted to maintain the degree of hydrogenation above 99 percent. Ultimately, the catalyst must be removed from the converters and regenerated. The hydrogen provided for the reaction in excess of requirements is recycled from the octane receiver to the converter, with the exception of the small exhaust gas stream. An increase in volume results from the hydrogenation reaction, the octane volumetric yield being 104 percent of the octene charge.

The guard tubes and converters are exchanger type reactors and a heat medium system comprising both a cold fluid and a hot fluid is provided. The cold fluid is used to remove the heat of reaction and maintain the optimum reaction temperature. The hot fluid circuit is used during startup of a converter and when forming fresh catalyst or reactivating spent catalyst. It also provides the heat medium for the feed vaporizer and hydrogen preheater. A hot oil system as shown is suitable.

Use of catalyst guard tubes is advantageous for sulfur removal and catalyst conservation. Ordinarily the guard catalyst is not effective in promoting hydrogenation itself. All the hydrogen feed may enter the octene vapors prior to passing through the guard tubes, but this large amount of hydrogen is not necessary for effective sulfur removal. Therefore, the major portion of the hydrogen feed may be introduced following the guard tube, thereby increasing the contact time of the octenes in the guard tube chamber. Roughly one volume of guard catalyst is sufficient for two volumes of active catalyst.

Hydrogen for plant operation should be free from sulfur, hence treatment before use may prove necessary if sulfur compounds are present. Convenient sources of hydrogen are the pyrolysis of natural gas and dissociation of ammonia. Some plants contain a separate catalyst tube for the hydrogen feed to convert traces of carbon monoxide to methane and to preclude formation of nickel carbonyl on cold spots in the catalyst bed; however, this is seldom necessary.

Deterioration of equipment by corrosion is negligible. Because of low temperature and pressure no special alloys or other materials are required.

Operation and control of hydrogenation plants are simple. The primary control point is the degree of unsaturation of the finished product as determined by titration with a weak brominating agent.

This process is licensed by Shell Development Company.

CATALYTIC HYDROGENATION PROCESS
COURTESY SHELL DEVELOPMENT COMPANY
HYDROGEN HEATER
VAPORIZER
CATALYST GUARD TUBE
CATALYST GUARD TUBE
CONVERTER
CONVERTER
CONVERTER
CONVERTER
COOLER
HOT OIL FURNACE
HOT OIL
COOLING OIL
RESIDUAL GAS
RECYCLE HYDROGEN COMPRESSOR
EXCHANGER
CONDENSER
OCTANE RECEIVER
OCTANE PRODUCT
OCTANE FEED
HYDROGEN
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

SMITHway Pressure Vessels for the Petroleum Industry

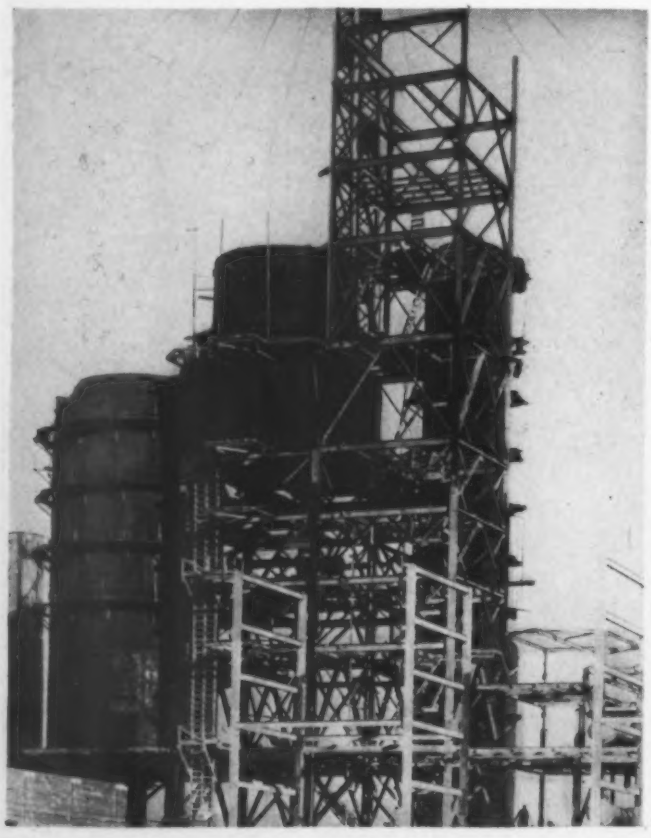
The world over, SMITHway vessels have withstood the test of time, temperature, pressure and corrosion. The first all-welded pressure vessel ever built was produced by A. O. Smith nearly three decades ago. After nearly two decades of continuous service the first SMITHway alloy-lined vessel is still in daily operation. Hundreds more of these vessels are proving themselves at temperatures ranging from 320° to 1050° F and even higher, and at pressures from vacuum to over 6,000 psi.

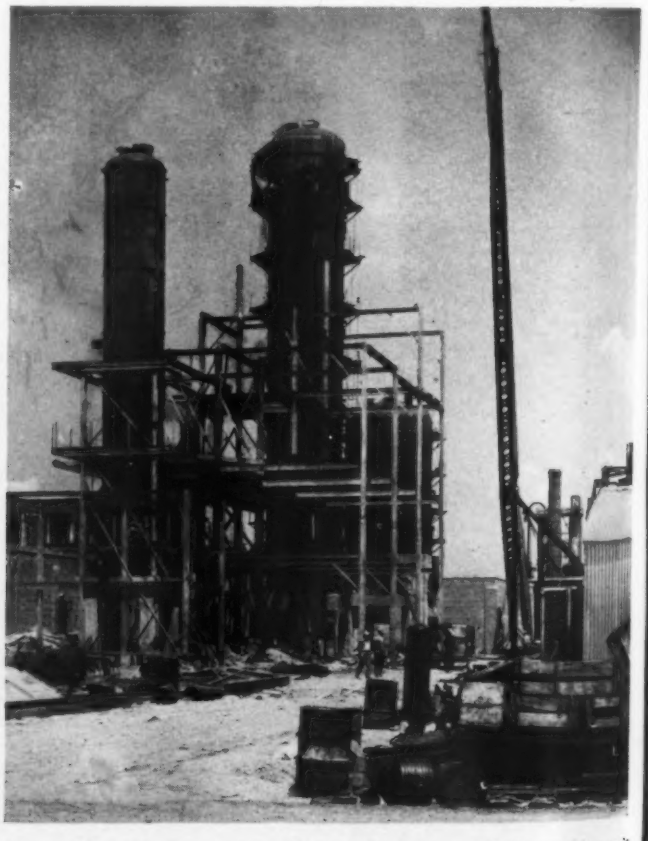
An alloy pre-testing program, based on performance of samples representative of fabricated alloys tested under actual operating conditions, coupled with over 25 years' accumulation of vessel manufacturing experience, provides SMITHway vessel users with a priceless fund of data on vessel design and construction.

Shipping limitations, plant locations, or building clearances no longer are bars to complete SMITHway service. A. O. Smith has developed a thoroughly proved technique for field assembly control and will guarantee vessels field assembled to A. O. Smith specifications under the quality control of an "on the job" Smith welding engineer.

Included as an essential part of all proposals and contracts is a set of SMITHway Vessel Specifications setting forth the responsibilities assumed by A. O. Smith.

Why not consult A. O. Smith, makers of all types of refinery vessels, on your vessel requirements?





A recent contract received by A. O. Smith called for manufacturing and field assembling the vessels for a high temperature catalytic cracking unit as well as a crude unit. On the left is the cat cracker consisting of a regenerator, reactor and main column. On the right is the crude unit consisting of fractionating and vacuum columns.

A. O. SMITH Corporation

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Gas Conversion Processes

THROUGH the development of several gas conversion processes light hydrocarbon gases which a few years back would have been discarded to plant fuel systems or to flares are now made into valuable products.

Early refiners made little use of gases released in distillation and cracking equipment. As gasoline production by thermal cracking increased, and as natural gasoline plants were built to recover light hydrocarbons, larger volumes of butanes and propanes were available, but generally were used to little advantage. After fractionation was substituted for weathering to set the vapor pressure of gasoline, butane assumed a rather stable position as a marginal component of gasoline. While a considerable market for liquefied petroleum gases gradually developed, and while some use was made of such material as feed for petrochemical manufacturing operations, still large excesses of butanes and lighter gases were wasted.

In the middle 1930's processes for the polymerization of butylenes and propylenes came into use. Refiners found that excess gases from cracking units could be converted readily into polymer gasoline of high blending octane value for use in motor fuels. Also, the poly plant product could be hydrogenated to make aviation gasoline stock. The equipment needed was comparatively simple and refineries in all capacity groups adopted the process.

Not much later, immediately before the second World War, a number of alkylation plants were placed in commercial operation. These plants utilized butylenes and isobutane as their feed stock and achieved the direct production of aviation fuel. The military requirements occasioned by the war led to the rapid construction of a great number of alkylation plants and such processes were by far the most important used for the manufacture of 100-octane aviation fuels and of the so-called super fuels.

Butylenes were essential also to the wartime synthetic rubber program and great volumes were used in the manufacture of both Buna and Butyl rubbers.

Because of these wartime demands butylenes and isobutanes, together with isopentane, became the most diligently pursued hydrocarbons in the history of the petroleum industry. To supplement short supplies of isobutane, several processes for isomerizing normal butane came into considerable use. In the same period isomerization processes were used also to some extent to convert heavier hydrocarbons to materials having higher octane ratings.

With the war emergency over, processes that were employed then without too much attention to economics were subjected to evaluation on their utility in normal commercial operations. The heavy demand for aviation gasoline dropped sharply and the synthetic rubber program was continued at greatly reduced volume.

Motor fuel demand, however, has greatly exceeded all expectations, and as a result, many process installations that were originally considered wartime measures have been continued in peacetime operation. While there is not the extreme shortage of butanes experienced earlier, many plants now are converting all available butylenes and propylenes into salable products. Sales of liquefied petroleum gases are at high levels so that there is something of a shortage of propane and heavier hydrocarbon gases. Petrochemical processing, too, is becoming a significant factor in the usage of light hydrocarbons.

One secondary but important effect of these developments has to do with the design of all kinds of process equipment. Fuel gas, which once included most of the butanes and lighter gases and was quite readily available in the average process plant, now may include only some of the propanes plus the lighter gases. In many plants fuel gas is a rather scarce and valuable utility. More care will be required to obtain good heat economy in process operations. Furnaces possibly will be more carefully designed to operate with smaller percentages of excess air, and heat exchangers will be more extensively used. In general, all practical improvements in heat efficiency will be given their due share of engineering attention and plant investment.

Catalytic Polymerization Process

UNIVERSAL OIL PRODUCTS COMPANY

THE UOP Catalytic Polymerization Process is one of the first of the many catalytic processes presently being employed in the petroleum refining industry. The development of the catalytic polymerization process offers to the petroleum refiner a means for converting a portion of fuel gas produced in cracking operations into a high grade motor gasoline, which increases the yield of gasoline obtainable from each barrel of crude oil processed. The process, in effect, then produces a finished gasoline of high value from a raw material which has the lowest value.

Most generally the process is used to polymerize propylene and butylenes found in the gases produced from thermal or catalytic cracking of topped crudes or gasoils. Propylene and butylenes can be polymerized either alone or in admixture, depending on the type of product desired.

Polymerization of the butylenes in the absence of other olefins can be controlled to yield a product which on hydrogenation will rate 90-95 octane number. This is referred to as a selective polymerization operation used to produce aviation gasoline components. Polymerization of propylene and butylenes in a mixture for the production of a fuel which will be blended in motor gasoline is referred to as a nonselective operation. This is the most common of all polymerization operations.

Propylene can also be polymerized separately to make either a motor polymer or a product rich in either C_6, C_9 or C_{12} polymer, depending on operating conditions. In recent years the butylenes normally found in refinery gases have in some cases become more valuable when converted to other products and consequently many plants are being operated for the polymerization of the residual propylene.

Briefly, the polymerization operation is one which consists of contacting the olefin-containing feed stream with the catalyst at temperatures in the range of 350° F. to 475° F. and at pressures ranging from 150 to 1200 psi. Processing conditions vary with the type of plant employed, with the feed stock processed and with the type of product desired.

There are two types of units presently being used for carrying on the process of polymerization. The most commonly used design is known as a chamber type. In this unit the catalyst is contained in a vertical chamber. Catalyst is loaded in one or more chambers, each containing multiple catalyst beds of 5 feet to 10 feet in depth.

Since the polymerization reaction is exothermic and the system in which the reaction takes place is essentially adiabatic, the temperature of the reactants increases as they travel through the catalyst beds. In order to maintain the catalyst temperatures in the range considered most effective for good olefin conversion and catalyst life, it is desirable to control the amount of reaction that occurs in the catalyst chamber by the recycling of hydrocarbons which have been through the catalyst chambers. This recycle material, being low in olefin content, will act to dilute the olefins in the fresh feed. It is possible in this manner effectively to limit the temperature rise through the reaction zone.

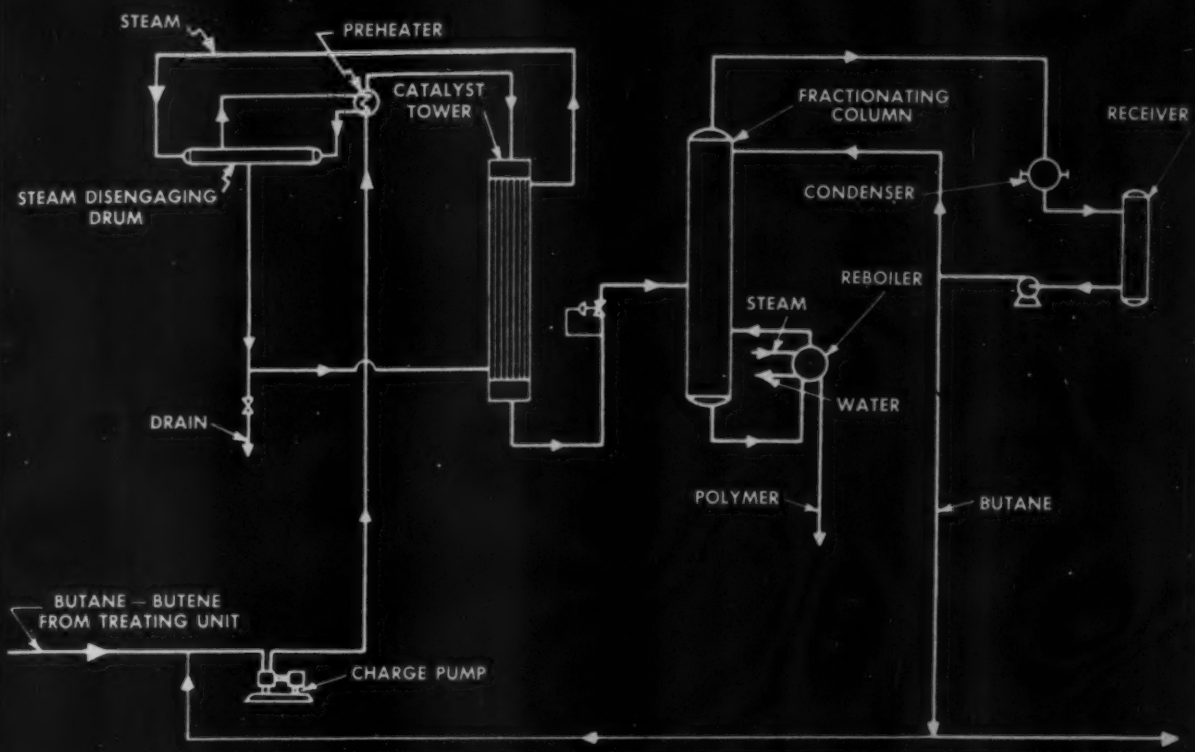
The other, or reactor type of unit, employs heat-exchanger type reactors. In this design the catalyst is loaded in vertical small diameter tubes which are surrounded with a cooling medium to remove the heat of reaction.

Reaction temperature control is therefore obtained by controlling the temperature of the cooling medium, and in this manner the reaction approaches isothermal conditions. In this unit it is therefore possible to process feeds of high olefin content without the use of recycle as in the case of the chamber type unit.

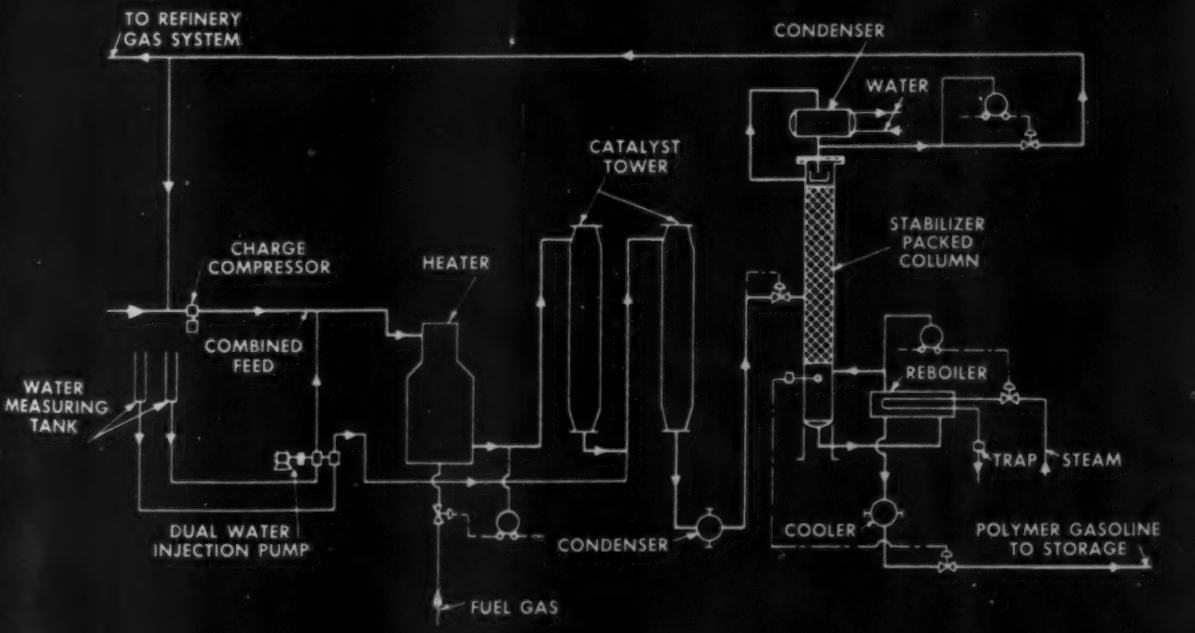
Reactor type units are most commonly used when the feed to be processed is high in olefin content such as is the case for C_3 and C_4 fractions produced from catalytic cracking.

Catalytic Polymer Properties

Type of Plants	Chamber	Reactor	Reactor
Pressure psig	500	1000	1000
Temperature °F.	370-465	400	450
Feed Stock	Cracking Plant Stabilizer Net Overhead Liquid	Butane-Butylene from Refinery Gas Concentration System	Propane-propylene from Refinery Gas Concentration System
Polymer Inspection:			
Gravity	64.0	64.3	62.4
IBP	96	78	144
10 Percent	156	152	204
30 Percent	210	238	244
50 Percent	235	258	266
70 Percent	279	....	284
90 Percent	366	379	330
EP	430	416	402
Rec.	97	95	....
RVP	9	11	4
Average Polymer Properties:			
Bromine No.	130-150		
Induction Period:			
Clear	15 min.		
+ 0.025 Percent UOP No. 1	500-550		
+ 0.005 Percent UOP No. 4	450-500		
Copper Dish Gum	25 mgs.		
ASTM Octane Number Clear		82.5	
ASTM Octane Number +3 cc TEL		85.0	
Research Octane Number Clear		97.0	
Research Octane Number +3 cc TEL		100.0	
Polymer Blending Values:			
5-10 Percent Blends in Average Straight Run Gasoline			110-120
5-10 Percent Blends in Thermal Cracked Gasoline			95-100

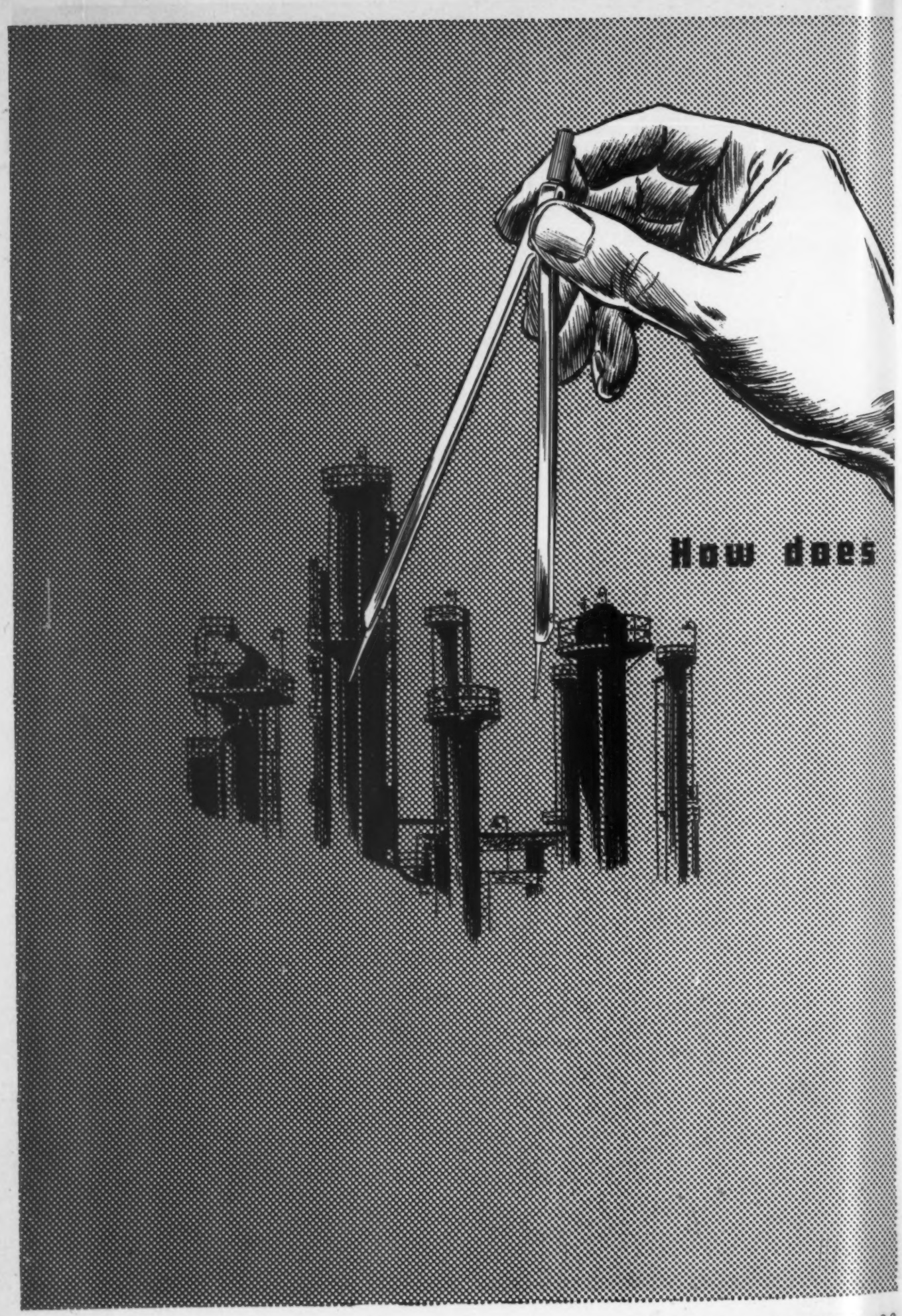


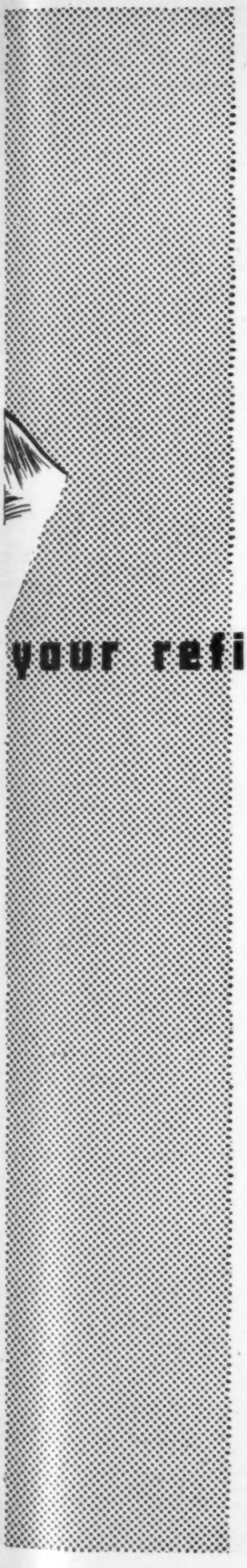
CATALYTIC POLYMERIZATION PROCESS—REACTOR TYPE



CATALYTIC POLYMERIZATION PROCESS—CHAMBER TYPE

COURTESY, UNIVERSAL OIL PRODUCTS COMPANY


How does



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Universal Service Protects Your Refinery

Catalytic Polymerization Process

THE M. W. KELLOGG COMPANY

CATALYTIC polymerization is a process for chemically uniting the propylenes and butylenes produced by thermal or catalytic cracking to produce a liquid which is an excellent component of motor and premium gasoline, and, after hydrogenation, of aviation gasoline.

Under normal circumstances the average refiner is able to include an appreciable amount of butanes in motor gasoline, but he usually has excess C_4's over and above this amount. Consequently, it is usually economic to polymerize propylene and butylene, and merely use the remaining saturated butanes for Reid-vapor-pressure fortification. Such catalytic polymerization moves propylenes and butylenes from fuel gas into the hydrocarbon group boiling in the gasoline range.

The catalyst employed in Kellogg units is copper pyrophosphate. It has a normal life of 100-150 gallons of polymer per pound of catalyst.

The liquid stream of C_3 and C_4 hydrocarbon reactants from the gas-recovery system is first treated to remove hydrogen sulfide and then washed with caustic to remove mercaptan sulfur. It is then heated to 300-400° F. and introduced into the reactor which is kept at about 900 psig to maintain the material in the dense phase.

After the reaction occurs, the reactor effluent proceeds to conventional stabilizing equipment as follows:

The hot product gases leaving the reactor exchange a portion of their heat to the incoming charge and then pass to a depropanizer where propane and any unconverted propylene is routed either to plant fuel or to bottled gas. The depropanized product flows to a debutanizer where butanes are removed, leaving the high-octane polymer gasoline as the bottom product. The butanes produced are used for motor-fuel-blending purposes.

Essentially there are two types of operation, depending on the feed stock.

(a) *Codimer Operation,* which is polymerization of butylenes only and which uses a hydrocarbon feed stock of essentially 100 percent C_4's. The operation usually is carried out at temperatures on the order of 300° F. The product, called codimer, is unsaturated, but when hydrogenated has a clear ASTM octane number on the order of 95 and is, therefore, a very acceptable component for aviation gasoline.

(b) *Motor Fuel Operation,* uses for a feed stock mixtures of C_3's and C_4's and the operation usually is carried out at approximately 375-400° F. The polymer is unsaturated and has a clear octane of about 83 ASTM and 96-98 CFRR. Its blending value, however, is considerably over 100 by either rating method when the percentage in the blend is not more than 10 or 15 percent.

A common type of reactor is the tubular reactor which is essentially a heat exchanger in which the tubes are filled with catalyst and surrounded by a bath of gasoil which is circulated to remove the heat of reaction. The reaction temperature is automatically controlled.

Typical Catalytic Polymerization Results

	Codimer Operation	Motor Fuel Operation
Conversion, percent of Olefin	80	90
Octane No.—ASTM Clear	95	83
Octane No.—CFRR Clear	....	96-98
RVP of Polymer, psi	2.0	2.0
IBP of Polymer, °F	148	170
EP of Polymer, °F	421	431
Gravity of Polymer, °API	60.4	60.1

Licenses are available from The M. W. Kellogg Company. Licenses are granted under the Polymerization Process Corporation patents, and patent rights of Phillips Petroleum Company, Standard Oil Company (Indiana), Standard Oil Company (New Jersey), The Texas Company, and The M. W. Kellogg Company.

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(a) The castings shall conform to the following requirements as to chemical composition:

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Chromium, per cent	25.0–28.0
Nickel, per cent	10.0–12.0
Manganese, per cent	0.50–1.25
Silicon, per cent	0.25–1.75
Sulphur, maximum per cent	0.05
Phosphorous, maximum per cent	0.05
Carbon	Within limits of Paragraph 4 (c)

(b) The chemical limits for all other added alloying elements, such as molybdenum, titanium and nitrogen, shall be specified by the manufacturer. The material shall conform therewith.

CREEP STRESS AND CERTIFICATE:

(a) The indicated creep rate of the material shall not exceed 1% in 10,000 hours when stressed by a tensile load not less than the following:

TEMP. °F.	STRESS PER SQ. IN. 1% CREEP IN 10,000 HOURS GRADE "F"
1400°	4200#
1500°	3700#
1600°	3200#
1700°	2600#
1800°	1950#

(b) Manufacturer shall submit standard creep data and residual physical properties made on material of his own manufacture and certified by an ACCREDITED LABORATORY. The chemical analysis of the creep test specimens shall be reported, including nitrogen content and any other special alloying elements.

TENSION TEST FOR DUCTILITY AFTER HEATING:

(a) Each heat shall conform to the following requirements as to tensile properties after the specimen has been subjected to a temperature of 1400° F. for 24 hours and furnace cooled to 400° F. Test shall be made at room temperature.

	GRADE "F"
Tensile Strength, min.lb./sq.in.	80,000
Yield Strength, min.lb./sq.in.	35,000
Elongation in 2", min. per cent	9.0

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MECO

Hot-Acid Polymerization Process

SHELL DEVELOPMENT COMPANY

THE SHELL COMPANIES have developed a method for producing isooctane from the butylenes present in refinery gas. Commercial size plants have given satisfactory performance over many years and the results in different refineries have proved the merit of this catalytic use of sulfuric acid. In this process, both normal butylene and isobutylene hydrocarbons react to produce isooctenes. The yield of octenes obtained from a typical butane-butylene fraction obtained from thermal cracking operations is in the order of twice the isobutylene content of the feed stock; however, this is to some degree dependent upon the normal butylene-isobutylene ratio in the charge. When these octenes are hydrogenated the resultant product has an octane number closely approaching that of pure isooctane. Feed stocks in different refineries are seldom the same, but the process has operated successfully on C_4 fractions varying widely in content of normal and isobutylene.

The flow diagram shows the major pieces of equipment involved in the hot-acid process. The charge of butane-butylene fraction enters the system continuously through the recycling line of the reactor time tank. The small fresh acid make-up requirement is also added to this recycling line. The mixture of recycled stock, fresh acid and butane-butylene fraction goes first to a reactor cooler and then to a reactor time tank where the polymer is produced. The reaction is exothermic but sufficient cooling is provided to maintain a temperature of about 175° F. This temperature promotes rapid polymerization so that the absorbed butylenes have only a transitory life in the acid phase, the polymer being immediately reabsorbed in the hydrocarbon phase. The process is thus truly catalytic. Present practice is to use a sulfuric acid concentration of approximately 70 percent and to provide from 10 to 15 minutes time of contact.

The mixture of hydrocarbon, polymer and acid leaving the reactor time tank flows through the recycle line into a reactor separator, in which the major portion of the acid is removed by gravity separation. The hydrocarbon and polymer with the remaining acid from the separator flows into a final acid settler. Then the hydrocarbon and polymer flow through a vaporizer to the debutanizer, where the butane and surplus normal butylene are removed overhead, and a bottom product, consisting of crude polymer, is withdrawn. A two percent soda solution is fed to this column also to neutralize any esters, which may decompose in the column reboiler, and to provide the neutralization reagent for the hydrolyzing step to follow.

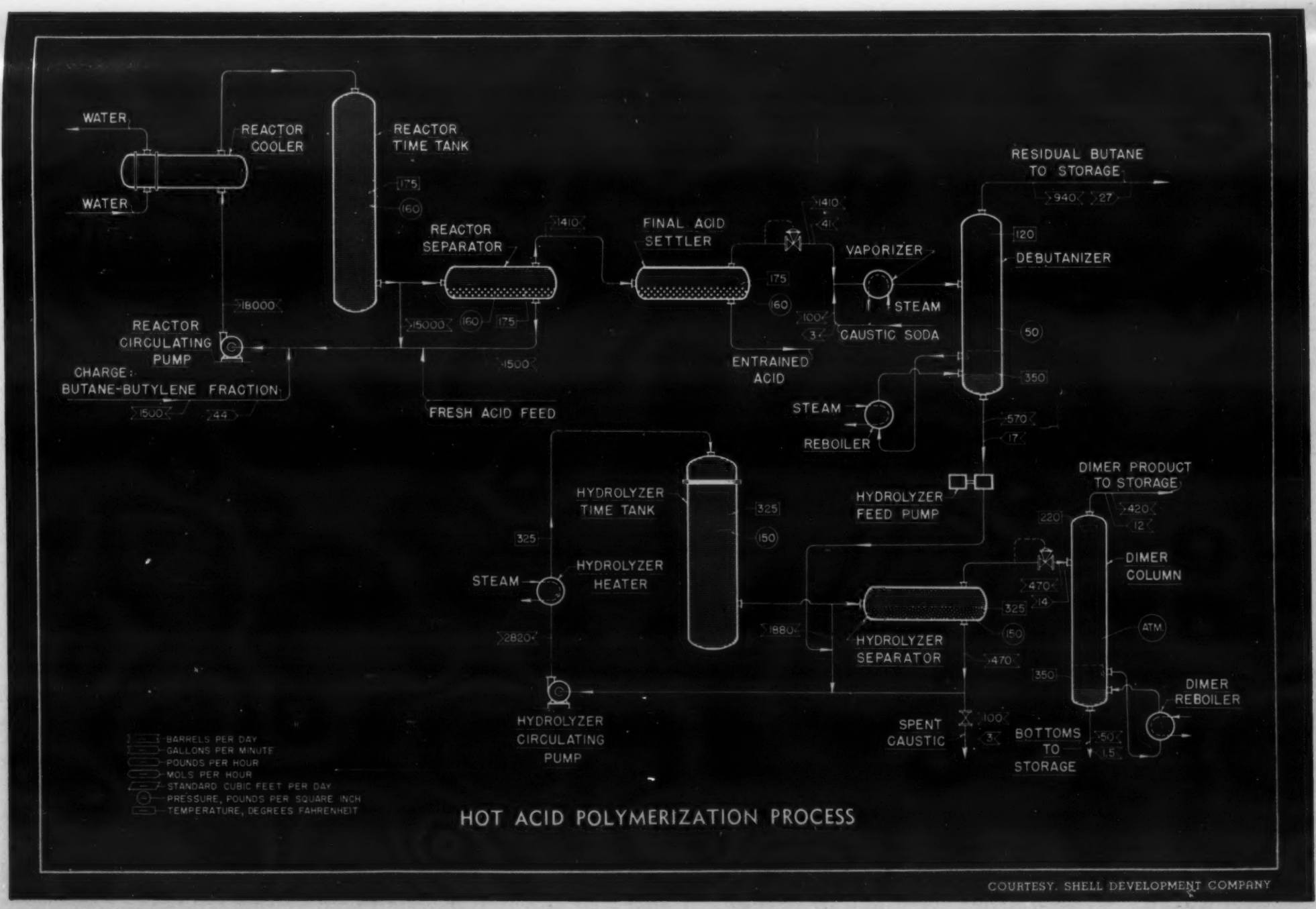
The soda solution and the polymer are delivered to the recycle line from the hydrolyzer time tank. The recycle line flows to a heater and thence to the hydrolyzer time tank in which all entrained acid is neutralized and acid sulfates are hydrolyzed with caustic. The caustic-polymer mixture then passes to a hydrolyzer separator, where the spent caustic is removed, the hydrocarbon and polymer residue being delivered without further heating to the dimer column. The dimer, which is the final product, is separated from the heavier hydrocarbon and delivered overhead to storage.

The chief advantage of this system is that normal butylene as well as isobutylene is used to produce the polymer (or dimer). Normal butylenes do not polymerize with themselves as readily as do isobutylenes, but in the presence of isobutylenes the normal butylenes will react readily to form interpolymer and it is for this reason that the yield of polymer is equivalent to twice the amount of isobutylene present in the charging stock. The interpolymer, passing into the hydrocarbon phase as rapidly as it is formed, is diluted by the inert butanes. Its activity is thus reduced, leaving little opportunity for the formation of trimers or rearrangement to polymers of lower octane rating. Consequently, the polymer formed is about 90 to 95 percent octenes. The product obtained upon hydrogenation has an octane number averaging between 98 and 99.

This process has also found advantageous application to production of a normal butylene rich stream for chemical process utilization. The operating conditions may be adjusted to effect virtually complete polymerization of the isobutylene while removing only a minor portion of the normal butylene. Hence, a butane stream rich in normal butylene uncontaminated with other unsaturates results. In this role the process serves to produce, firstly, an octane dimer which upon hydrogenation becomes a high octane blending constituent, and, secondly, a normal butylene rich stream suitable for chemical processing.

Operation of the unit is simple; one operator per shift is sufficient. Both polymerization and neutralization are automatic once the rate of flow is established.

This process is licensed by Shell Development Company.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STEAM
DIMER COLUMN
HYDROLYZER SEPARATOR
HYDROLYZER CIRCULATING PUMP
SPENT CAUSTIC
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DIMER REBOILER
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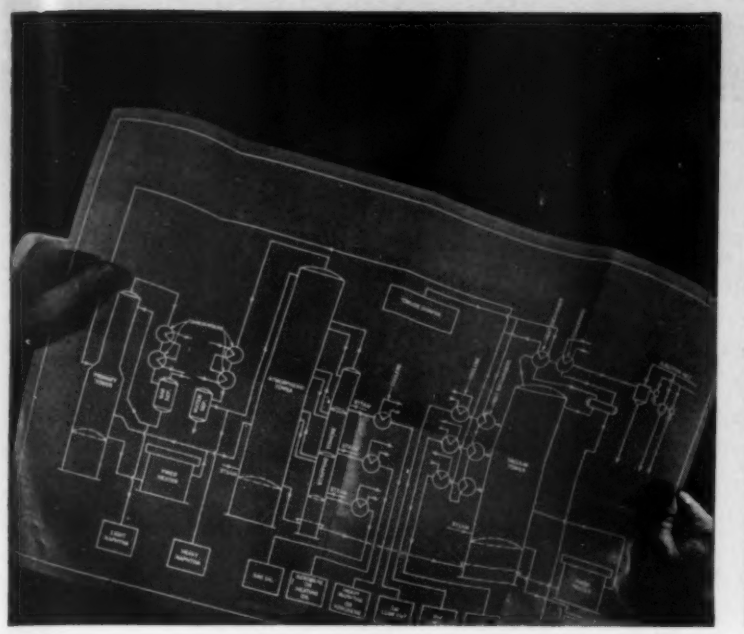

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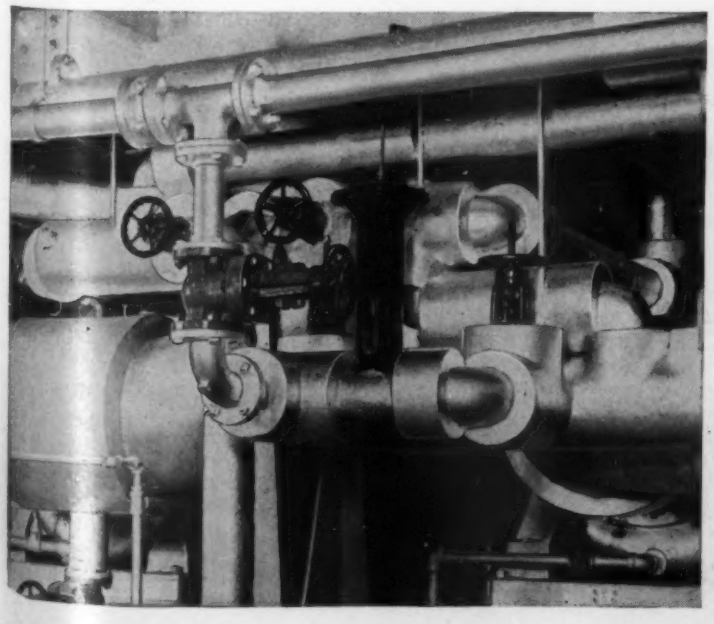

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Catalytic Alkylation Process

THE M. W. KELLOGG COMPANY

ALKYLATION as practiced by the petroleum industry is the catalytic method of combining an isoparaffin (usually isobutane) with an olefin (propylene, butylene or amylene) to produce saturated isocompounds. The product of the reaction—alkylate—has a high-octane rating and is therefore employed as a blending agent in aviation and motor gasoline. It also has high heat value, giving it more power than other hydrocarbon fuels.

The catalyst used is either sulfuric acid or hydrofluoric acid. Yields and quality of products with either catalyst are essentially the same. The majority of the installations, however, employ the lower-cost sulfuric acid since it is less corrosive and the spent acid can be re-used in other refinery treating operations.

In the Kellogg flow sheet, the olefin is dissolved in concentrated sulphuric acid and violently agitated with an excess of isobutane at a temperature of 35-45° F. and a pressure several pounds above atmospheric. Under these conditions, the reaction proceeds readily.

Since the reaction is exothermic, the heat of reaction is removed by auto-refrigeration. Refrigeration is also required to remove the heat imparted by the emulsion pumps. Conditions are maintained at the bubble point of the mixture and a sufficient amount of hydrocarbon is allowed to vaporize to cool the mixture to the control temperature. The vaporized hydrocarbon is compressed, condensed and returned to the system.

The product of the reaction is separated from acid in the center section of the combination reactor-settler and after caustic treatment is pumped to the isobutane tower.

The isobutane overhead from this tower is returned to the process while the alkylate bottoms containing any normal butane present in the original feed is sent to a debutanizer tower where butane is removed as an overhead product. The alkylate bottoms usually are separated in a rerun tower into a light aviation cut and a heavy alkylate.

The major factors affecting product quality and other operating results are:

1. Reaction Temperature
2. Olefin Concentration
3. Isobutane Concentration
4. Acid Strength

Each of these factors will be discussed briefly.

1. Reaction Temperature

The lower the temperature at which the reaction is carried out, the better the quality of the alkylate and the lower the acid consumption. On the other hand, below 35° F. the sulfuric acid-hydrocarbon emulsion tends to get so stiff as to make operations difficult.

2. Olefin Concentration

The lower the concentration of olefin is kept at any time, the lower is the tendency toward polymerization and the higher the tendency toward alkylation. This is effected by circulating large volumes of emulsion as compared with the amount of material in the reactor, and controlling the addition of olefins so as to maintain this low concentration.

3. Isobutane Concentration

Normal butane does not enter into the reaction and is a harmful diluent in the sense that it reduces the chances for contact between the isobutane and the olefin-bearing acid. It is, therefore, advisable to keep the isobutane concentration in the reactor as high as possible.

4. Acid Strength

While 98 percent is the strength of the acid usually added to the process, a more important factor is the average strength of the acid contained in the reactor. In the case of butylenes and propylenes, this strength should not be less than 90 percent. In the case of amylenes, the acid strength can drop to 85 percent without seriously affecting the quality of the alkylate.

Most of the alkylation units in this country were designed for operation on butylenes but today many units are operating on propylenes. Butylenes give the best product quality and the lowest acid consumption. The same units also can be used for alkylating amylenes. Typical results are given in the following table.

Typical H_2SO_4 Alkylation Results

	Butylenes	Propylene	Amylenes
Alkylate—Volume Percent on Olefin	170	170	160
Isobutane—Volume Percent on Olefin	110	125	110-120
Acid Consumption—Lb. per gal. Alkylate	0.7	3	1-2
ASTM Octane rating of Alkylate	94	89	90
Reid Vapor Pressure of Alkylate—lb./sq. in.	3-4	5	2
End Point of Alkylate—°F.	350	350	350

Because the sulfuric acid is of high concentration and the plants are operated at below atmospheric temperature, corrosion is negligible and steel equipment can be used throughout. Individual requirements vary widely and the economics depend largely on the availability of the raw materials at low cost.

The M. W. Kellogg Company, New York, has been appointed licensing agent for the process for the United States and foreign countries. Licenses granted are under the patents and patent rights of Texaco Development Corporation, Standard Oil Development Company, Shell Development Company, and Anglo-Iranian Oil Company, Ltd.

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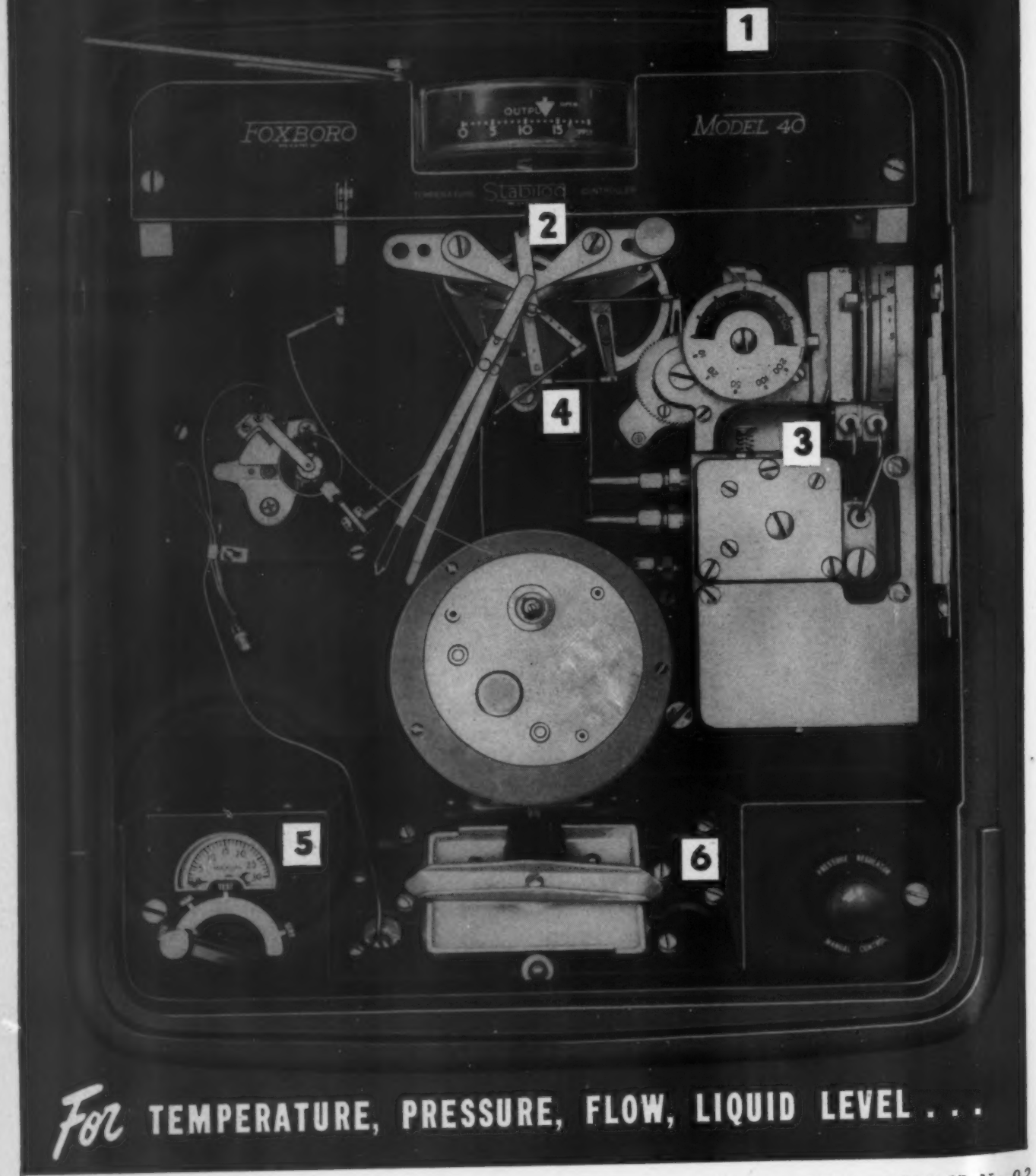

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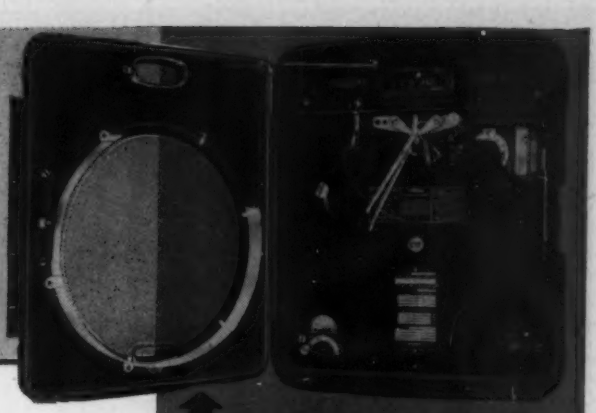

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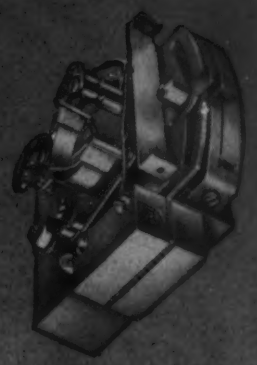

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Paraffin Alkylation Processes

UNIVERSAL OIL PRODUCTS COMPANY

ALKYLATION of paraffins is a reaction in which an isoparaffin is combined with an olefin to produce a branched chain paraffin hydrocarbon of higher molecular weight.

The alkylate product boils in the gasoline range, and has an A.S.T.M. octane number of 88-95 depending on the conditions under which it is made and the character of the feed. Alkylate has a high lead susceptibility making it an ideal blending component.

Extensive use was made of the alkylation process during the war to produce a high-octane-number component of aviation gasoline. Although initially considered as a high-cost operation, many simplifications have been achieved in the process which make it attractive for producing a high octane gasoline blending component for improved quality motor fuels.

The operation of this process is particularly attractive in those cases where low cost isobutane is available in sufficient quantities to alkylate with available olefins. Although some isobutane is present normally in admixture with olefins in gases produced from thermal cracking operations, it is not available in sufficient quantities to alkylate with the available olefins. In these cases isobutane must be made up from other sources such as separation from natural gas or isomerization of available normal butane.

Light gases from catalytic cracking will in some cases, depending upon the conversion and type of catalyst employed, contain a relatively high percentage of isobutane, sufficient in quantity to alkylate with a substantial portion of the butylenes and propylenes available. In such instances the alkylation process is well suited to operate in conjunction with catalytic cracking to obtain the maximum yield of high octane gasoline.

The UOP Paraffin Alkylation Process is a catalytic process utilizing either hydrofluoric acid or sulfuric acid as the catalyst. The materials which enter into the alkylation reaction are isobutane and olefins. The olefins ordinarily employed are butylene, propylene, and amylene.

When isobutane is alkylated with butylenes, the volume of isobutane should be approximately 20 percent greater than the volume of olefin, yielding 160 percent by volume of rerun alkylate.The yield of alkylate and the consumption of isobutane will vary slightly, depending on operating conditions and the olefin charged.

In order to favor the alkylation reaction and minimize side reactions such as polymerization, the feed to the alkylation reactor is provided with an excess of isobutane over and above that required for alkylating the olefins. This is accomplished by recycling a stream rich in isobutane. The higher the ratio of isobutane to olefin in the feed to the reaction zone, the higher the yield and quality of the alkylate produced. Economic considerations and product quality requirements will determine the ratio used. For most cases the recycle of isobutane should be such as to provide five to ten times as much isobutane as olefin in the total feed to the reactor.

The alkylation reaction is carried out at temperatures in the range of 45° F. to 100° F. and at sufficient pressure to maintain the reaction mixture in a liquid phase. The lower temperatures are employed when utilizing sulfuric acid. This relation does not apply when hydrofluoric acid is used; therefore, the reaction may be carried at the higher temperature when using this catalyst. The operating pressure is such as to maintain the reaction mixture in a liquid phase.

The reactor system used for this process provides for the intimate mixing of the dry hydrocarbon feed with the catalyst as well as cooling of the reaction mixture to remove the heat liberated by the reaction.

After completion of the reaction in the reaction zone, it is necessary to separate the total hydrocarbon reaction mixture from the catalyst as well as to separate the alkylate product from isobutane recycle and the hydrocarbons, such as normal butane and propane, which do not enter into the reaction.

Because of the difference in physical characteristics of the hydrofluoric acid and sulfuric acid catalysts which may be used for promoting the reaction, the design of the unit will differ somewhat in the methods for contacting and separating the catalyst from the hydrocarbon mixture, depending on the catalyst used.

In the alkylation process one of the important operating variables is the strength of the acid. Since the acid catalyst in service becomes contaminated with dissolved tars, it is necessary to continuously replace a portion of the acid catalyst with fresh material to maintain a desired level of acidity. When sulfuric acid is used as the catalyst the acid strength is maintained by introducing fresh high strength acid and discarding spent acid, which can be used for other operations in the refinery, such as gasoline treating. Hydrogen fluoride can be purified by a distillation step to effect the separation from the heavy hydrocarbon contaminants. The hydrofluoric acid purification is carried out on a continuous basis, returning the purified acid to the reaction zone to maintain the desired acid strength.

Catalytic Alkylation—Typical Results

FEED STOCKS	Butylenes from Catalytic Cracking	Amylenes from Catalytic Cracking	Propylene
Operating Conditions:			
Isobutane-Olefin Ratio	5:1	10:1	10:1
Reaction Temp.—°F. HF	90	60	90
Reaction Temp.—°F. H_2SO_4	45	45	45
Acid-Hydrocarbon Ratio	1:1	1:1	1:1
Acid Strength	87-90	87-90	87-90
Product Quality:			
Octane Number ASTM Clear	92	89	88
E.P.	350	350	350
RVP	4.5	2.3	2.0
Conversion Factors:			
Vol. 350° EP Alky./Vol. Olefins	1.6	1.6	1.6
Vol. 350° EP Alky./Vol. Isobutane	1.25	1.1	1.1
Catalyst Consumption:			
Using HF as catalyst*—Gallons Alkylate/lb Acid	85-150	60-100	60-100
Using H_2SO_4 as Catalyst—Gallons Alkylate/lb Acid	2.0	0.5	0.5

* One of our plants operating at 5000 barrels or more per stream day, has averaged 168 gallons of alkylate per pound of acid for an entire year's operation.

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SPENT CAUSTIC
PROPANE
BUTANE
ALKYLATE
NOTE: — DEPROPANIZER USED ONLY WHEN FEED CONTAINS PROPANE-PROPYLENE FRACTION
HYDROGEN FLUORIDE ALKYLATION PROCESS
DE-ISOBUTANIZER
DEPROPANIZER
DEBUTANIZER
CONTACTOR
WATER COOLING
ACID SETTLER
ACID REGENERATOR
TAR
ISOBUTANE AND HF RECYCLE
BAUXITE TOWERS
HF STRIPPER
BUTANE
ALKYLATE
PROPANE TO STORAGE
OLEFIN FEED
ISOBUTANE
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LIQUID LEVEL GAGES: Reflex, Transparent, Tubular, Distant Reading, Heated, Cooled, Vacuum Jacketed, Non-Frosting and High Pressure. LIQUID LEVEL GAGE VALVES: Standard, Offset and Jacketed. WATER COLUMNS, TRI-COCKS, GAGE ILLUMINATORS, SIGHT GLASSES, SPECIAL DESIGN Sight Glasses, Visual Retorts, etc., made to fit individual requirements.

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GAGE ILLUMINATORS. New wedge light principle with Plexiglass construction assures even diffusion of light through gage glass. Available in explosion proof construction.

Reaction Chamber Gage

Gage Illuminator installed on back of Trulevel Distant Reading Boiler Water Gage

REACTION CHAMBER GAGES. With this unit the gage chamber acts as a reaction chamber which is capable of withstanding pressures up to 1,000 lbs. The charge can be agitated to test mixing qualities under various pressure and temperature conditions, before and after samples being taken from the outlets. A great many Jerguson

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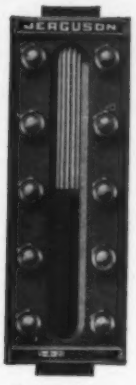
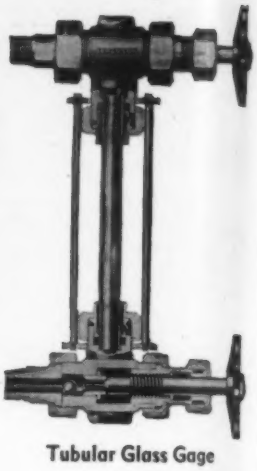
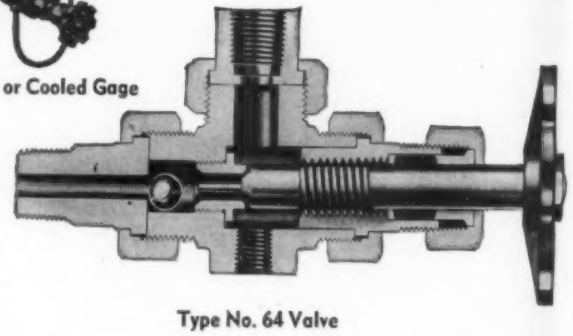


Reflex Gage



Tubular Glass Gage



Type No. 64 Valve

Reaction Chamber Gages have been designed and built for special applications. We shall be glad to work with you to design and build Reaction Chamber Gages to meet your needs.

Non-Frosting Gage

NON-FROSTING GAGES. These gages have special frost preventing unit in direct contact with gage glass. The unit protrudes beyond cover so that frost cannot build up across slot in cover forging. Ideal for plants in chemical or petroleum industries where low temperature fluids must be gaged.

FLANGED END GAGES. Where safety measures require the elimination of all threaded connections, all standard Jerguson Reflex and Transparent Gages, as well as valves, are available with flanged end connections. Gages may have flanged top, side or back connection.

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Ask for detailed Data Units on any Jerguson products, or write to us for engineering suggestions and assistance in the application of standard Jerguson gages and valves or the development of special gages and valves to meet your problems.

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This includes eleven standard tube alloys—as well as duplex tubes for special requirements. Plates for tube sheets are made in widths up to 156″ (circles to 160″). Plates and circles are accurately alloyed, commercially flat, true to dimensions, uniformly machinable, and unsurpassed in soundness and close-grained structure.

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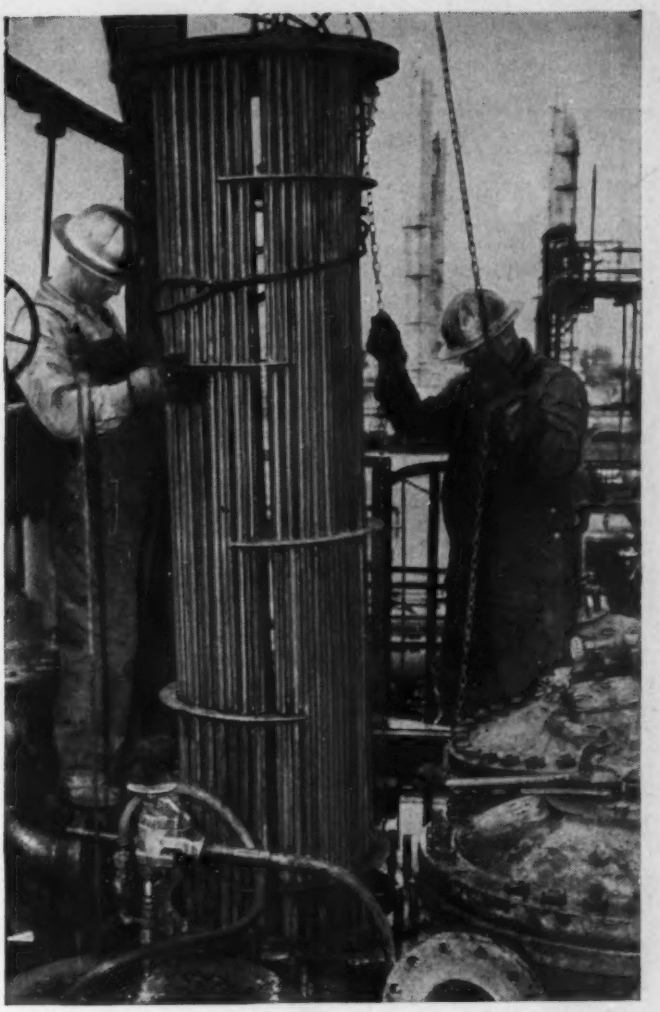
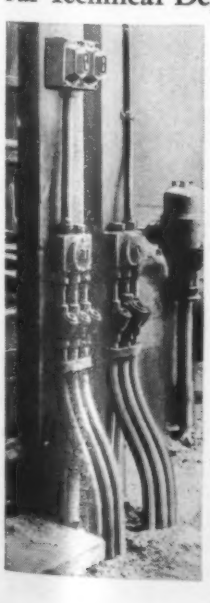


Installation of tube bundle equipped with Anaconda Arsenical Admiralty Condenser Tubes at the Houston, Texas refinery of the Shell Oil Company.

STANDARD ANACONDA CONDENSER TUBE ALLOYS

Arsenical Admiralty	Cupro-Nickel
Ambraloy (Aluminum Brass)	*Ambrac
Ambraloy (Aluminum Bronze)	Red Brass (85%)
Super-Nickel	Muntz Metal
	Deoxidized Arsenical Copper
	Deoxidized Copper

*Reg. U. S. Pat. Off.



Concentration of Everdur Electrical Conduit and Everdur fittings, located in a moist alcove between cooling towers at the Torrance Refinery of General Petroleum Corporation. After nine years in this corrosive atmosphere, the installation shows no sign of corrosion — while rustable conduit in the same area failed in a few months.

Other Anaconda Products

The American Brass Company produces a long list of coppers and copper alloys in various commercial forms (sheet, strip, plate, wire, rod, seamless pipe and tube, extruded and drawn special shapes and die pressed forgings). For information on all Anaconda Metal Products, write for Publication B-28.

ANACONDA
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Subsidiary of Anaconda Copper Mining Company

In Canada: ANACONDA AMERICAN BRASS LTD., *New Toronto, Ont.*

HF Alkylation Process

PHILLIPS PETROLEUM COMPANY

THE alkylation of an isoparaffin such as isobutane, with olefins, such as propylene, butylenes or amylenes, using hydrofluoric acid as a catalyst is a well-established process for producing high-octane-number premium aviation and automotive fuels.

The essential feature of the Perco HF Alkylation Process is the reaction of the olefins with an isoparaffin using liquid hydrofluoric acid as a catalyst. As shown in the flow sheet, liquid olefin-containing feed and isobutane usually are fed into the bottom of a mechanically-agitated reactor while liquid hydrofluoric acid catalyst is fed into the top. After intimate contacting in the reactor, the hydrocarbon and acid are separated in a settler. Acid is recycled to the contractor and the hydrocarbon is purified and separated by fractionation and other means.

The process is operated at ordinary atmospheric temperatures and at sufficient pressure to keep the hydrocarbon mixture liquid. The acid-hydrocarbon ratio most frequently used is in the neighborhood of 1:1. A large excess of isobutane is used over that required for reaction with the olefin. External isobutane-to-olefin ratios from 4 to 1 up to 12 to 1 have been used. The acid strength normally is maintained in the range of 85-95 percent hydrofluoric acid.

Several subsidiary operations are necessary to the process. The hydrofluoric acid catalyst is continuously purified by fractionation of a small portion of the circulating stream in an acid rerun unit as shown in the diagram. The hydrocarbon effluent is first treated to remove dissolved acid by fractionation in the azeotropic tower. The excess isobutane is removed by fractionation in a deisobutanizer, and the isobutane is separated from propane in a depropanizer before being recycled to the contactors. The deisobutanizer bottoms, containing all hydrocarbons heavier than isobutane, are catalytically defluorinated before being further fractionated. The fractionation steps following are dependent on the nature of the products desired from the process. In the flow sheet shown, normal butane, iso- and normal-pentane, aviation and heavy alkylate are all separated by fractionation.

Typical operating and other data are given in the accompanying tables. Certain features to be considered in designing a unit for peacetime motor and aviation fuel production would include:

1. Combination of the main azeotropic fractionation step with the deisobutanizing step, one fractionator serving both functions. A small azeo tower could be used on a bleed stream to a depropanization step. This change would result in savings in capital investment, utilities, and in maintenance.

2. Simplification of the acid-rerun equipment by use of a single column, acid-soluble oil and water being removed at the bottom of this column together, and by the use of a higher column pressure to eliminate the necessity of overhead-product pumps with attendant savings in maintenance.

3. In production of alkylate for motor-fuel blending, the fractionation of aviation alkylate from heavy alkylate could be eliminated in most cases with further savings in utilities and in capital investment for a new plant.

A very comprehensive treatise of the subject of HF Alkylation was published in a book by the same name by Phillips Petroleum Company, Perco Division, Bartlesville, Oklahoma, licensors of HF Alkylation.

TABLE 1

Typical Operation of HF Alkylation Unit

Charge to Unit:	Bbl. per Day
Propylene	89
Butylene	889
Amylenes	246
Total Olefins	1,224
Saturated Hydrocarbons*	
Total Feed	3,086
Recycle Isobutane, Bbl. per Calendar Day†	10,300
Yields:	**Bbl. per Day**
Aviation alkylate	2,070
Heavy alkylate	134
Isopentane	190
Normal Pentane	349
Propane	106
Acid-soluble oils	20
Aviation Alkylate Yield, bbl. per bbl. Olefin	1.7
Isobutane consumption, bbl. per bbl. Olefin	1.2
Operating Conditions:	
Isobutane-Olefin ratio (external)	7.5
Acid-Hydrocarbon ratio	1.0
Pressure, psi	150
Temperature, °F	102
Contact time, minutes	15
Acid Catalyst:	
Titrable acidity, Wt. Percent	89.6
Oils plus solids, Wt. Percent	3.0
Water, Wt. Percent	1.6

* Contains 418 bbl. Isobutane.
† Total stream—80 percent purity.

TABLE 2

Typical Properties of HF Alkylate

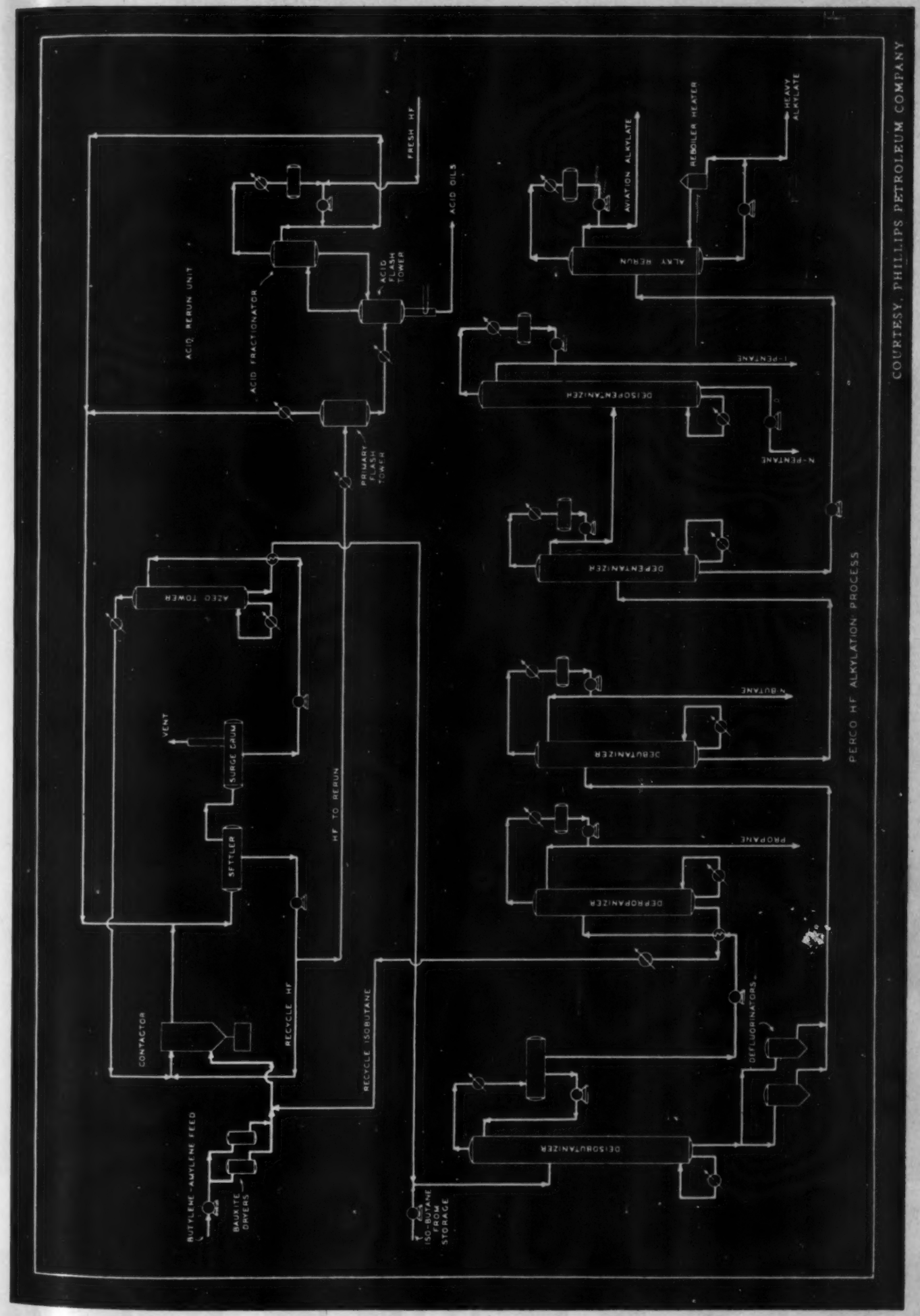
	Depentanized Aviation Alkylate	Heavy Alkylate
Gravity, °API	69.7	45.8
ASTM Distillation, °F.:		
IPB	178	416
10 Percent	200	428
50 Percent	220	448
90 Percent	239	539
End Point	327	616
Reid Vapor Pressure, lb.	2.3	....
Octane Number, ASTM	89.8	75
+1 cc TEL	100.7	
Octane Number, Research	91.6	
+ 1 cc TEL	99.7	

TABLE 3

Utility and Chemical Consumption HF Alkylation

Basis—3000 bbl. per day aviation alkylate was operation. 3400 bbl. per day motor alkylate-redesigned peacetime plant.

UTILITIES	Wartime Operation	New Design
Steam, lb. per hour	115,000	90,000
Electricity, kw	1,300	950
Water Circulation, gpm	5,300	4,600
Water, makeup, gpm	283	247
Fuel gas, cu. ft. per hr.	12,000	1,200
Chemical, lb. per month:		
Caustic Soda	9,850	9,850
Bauxite	105,000	32,000
Hydrofluoric Acid	60,000	27,000
Lime	1,500	1,500
Labor:		
Operation, man-hours per day	157	157
Maintenance, man-hour per day	105	105


PERCO HF ALKYLATION PROCESS
COURTESY, PHILLIPS PETROLEUM COMPANY
BUTYLENE - AMYLENE FEED
BAUXITE DRYERS
ISO-BUTANE FROM STORAGE
CONTACTOR
SETTLER
VENT
SURGE DRUM
AZEO TOWER
RECYCLE HF
RECYCLE ISOBUTANE
HF TO RERUN
PRIMARY FLASH TOWER
ACID FRACTIONATOR
ACID RERUN UNIT
ACID FLASH TOWER
FRESH HF
ACID OILS
DEISOBUTANIZER
DEFLUORINATORS
DEPROPANIZER
PROPANE
DEBUTANIZER
N-BUTANE
DEPENTANIZER
DEISOPENTANIZER
N-PENTANE
I-PENTANE
ALKY RERUN
REBOILER HEATER
AVIATION ALKYLATE
HEAVY ALKYLATE

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Butane Isomerization Process

SHELL DEVELOPMENT COMPANY

THE greatly expanded alkylation capacity, brought about over the past five years by the unprecedented wartime demand for high-octane aviation gasoline, has been largely maintained in operation to meet peacetime upgrading of the octane number of motor fuels. This postwar production of alkylate requires a high consumption of isobutane, used in its manufacture, which in many refineries is greater than the amount normally available from natural gasoline and cracking operations. The Shell Butane Isomerization Process provides means for supplying this deficiency.

Briefly, the process is one in which normal butane is converted, under the influence of a catalyst, to isobutane. In many naturally occurring mixtures of butanes, or synthetically prepared cuts of butylenes and butanes, the proportion of normal butane is greater than that of isobutane.

A typical analysis of a C_4 cut from reforming operations (i.e., the fraction containing butanes and butylenes) is as follows:

	Percent by Weight
Isobutylene	20.0
Normal butylenes	27.0
Isobutane	18.5
Normal butane	34.5
	100.0

Such a mixture is not suitable as the sole charge stock for alkylation operations as it is deficient in isobutane. However, if normal butane, which usually is discarded from the operation, is taken as charge stock for an isomerization unit, the requisite quantity of isobutane will be formed for charging to an alkylation unit. Some of the normal butane charged to the isomerization unit is unconverted and accompanies the isobutane. The unconverted normal butane is segregated in the fractionation section of the alkylation plant and is once more charged to the isomerization unit, thus the normal butane is recycled through isomerization and alkylation plants until conversion is completed, eliminating or materially reducing the necessity for obtaining natural isobutane.

The process consists of passing normal butane in vapor form, at a moderate temperature, through one or several reactors, each containing a granular bed of catalyst, a constituent of which is aluminum chloride. Dry hydrogen chloride gas is incorporated with the butane vapors to act as a catalyst promoter. The construction of the reactors may provide for a single catalyst bed, or may be of a heat-exchanger type with the catalyst placed in tubes surrounded by circulating cooling liquid as shown on the accompanying diagram.

The discharge from a single pass through the reactors flows to a receiver through a cooler and a refrigerated condenser which condenses all the butanes, the hydrogen chloride going into solution. A small amount of uncondensable gas is vented from the receiver after being scrubbed with product recycle to recover hydrogen chloride, and the condensate is charged to an HCl stripping column, which effects substantially complete removal of recycle hydrogen chloride gas.

The accompanying flow sheet shows the essential features of a typical isomerization plant. The hydrocarbon charge is picked up by the charging pump and passed through a drying step which extracts the last trace of moisture. The feed is completely vaporized and the flow is divided in the desired proportions to the reactors; each individual stream receives its correct amount of hydrogen chloride recycle gas.

The vapors pass through the catalyst beds and exit streams combine before flowing through coolers and condensers. The condensate is run to a receiver from which it is pumped to the HCl stripping column. Butanes, substantially free of hydrogen chloride, flow from the base of the column, while vapors, consisting largely of hydrogen chloride, are recycled to the reactors. The last traces of hydrogen chloride in the isomerized product are removed by caustic scrubbing.

No special materials are required for the construction of the plant since dry hydrogen chloride, under the conditions prevailing in the process, is not corrosive. Temperatures and pressures under which the plant operates are moderate, consequently the required capital outlay is low.

The catalyst has been developed specially for the purpose. It is rugged, cheaply produced, and has long life; these factors combined make its regeneration unnecessary. The catalyst is resistant to poisons, but it is necessary to reduce the proportion of olefins and pentanes in the charge stock to a low figure, as both of these produce objectionable by-products and tend to shorten the life of the catalyst.

The process is licensed by Shell Development Company, New York.

BUTANE ISOMERIZATION PROCESS
1000 BARRELS PER DAY PRODUCTION OF ISOBUTANE
HCl MAKEUP
REACTORS
CIRCULATING OIL COOLERS
N-BUTANE FEED
CHARGE PUMP
FEED VAPORIZER
Steam
REACTOR GUARD SECTION
HOT GAS
COOLING OIL PUMP
EXCHANGER
DRYER
DRYER
COOLER
MIXER
HCl RECYCLE
VENT GAS SCRUBBER
GAS VENT
CONDENSER
REFRIG. CONDENSER
ACCUMULATOR
STRIPPER CHARGE PUMP
HCl STRIPPER
REBOILER
Steam
NEUTRALIZED PRODUCT TO FRACTIONATION
CAUSTIC WASH
CAUSTIC CIRCULATING PUMP
ISOBUTANE
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
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Pentane and Butane Isomerization Processes

STANDARD OIL COMPANY (INDIANA)

THE Indiana Isopentane Process is a method for isomerizing normal pentane to isopentane. In this process, as in the Indiana Isomate Process, the catalyst is a liquid aluminum chloride-hydrocarbon complex promoted with anhydrous hydrogen chloride. In the Isopentane Process, cracking and disproportionation are suppressed by inclusion of a small amount of cyclic hydrocarbons in the feed to the isomerization unit. Although both naphthenes and aromatics are suitable for this purpose, benzene is most effective. From about 0.3 to 1.3 weight-percent based on charge is sufficient to inhibit cracking and to maintain the catalyst activity. As in the Isomate Process, conversion is accomplished by passing the hydrocarbon solution of hydrogen chloride in a dispersed phase upward through the catalyst complex. Suitable operating conditions for pentane isomerization include a reactor temperature of 195-225° F. and a reactor pressure of 275-325 pounds per square inch.

The use of benzene to replace hydrogen as a cracking inhibitor permits operation at a lower pressure and effects a considerable reduction in initial investment. The process is characterized by high yields and low catalyst consumption.

An Indiana Isopentane Process installation has the added advantage that it can with little change be used for isomerization of normal butane. With such operations, a cracking inhibitor is unnecessary, since normal butane can be isomerized with negligible tendency toward cracking or disproportionation.

The Isopentane Process usually is operated on a recycle basis. An installation has been operated at the Texas City refinery of the Pan American Refining Corporation in conjunction with an alkylation unit, in the manner shown in the diagram. Deisobutanized alkylate is charged to a depentanizer, here the pentanes are taken overhead and subsequently split in an isopentane tower. The feed-preparation towers also are used for product fractionation.

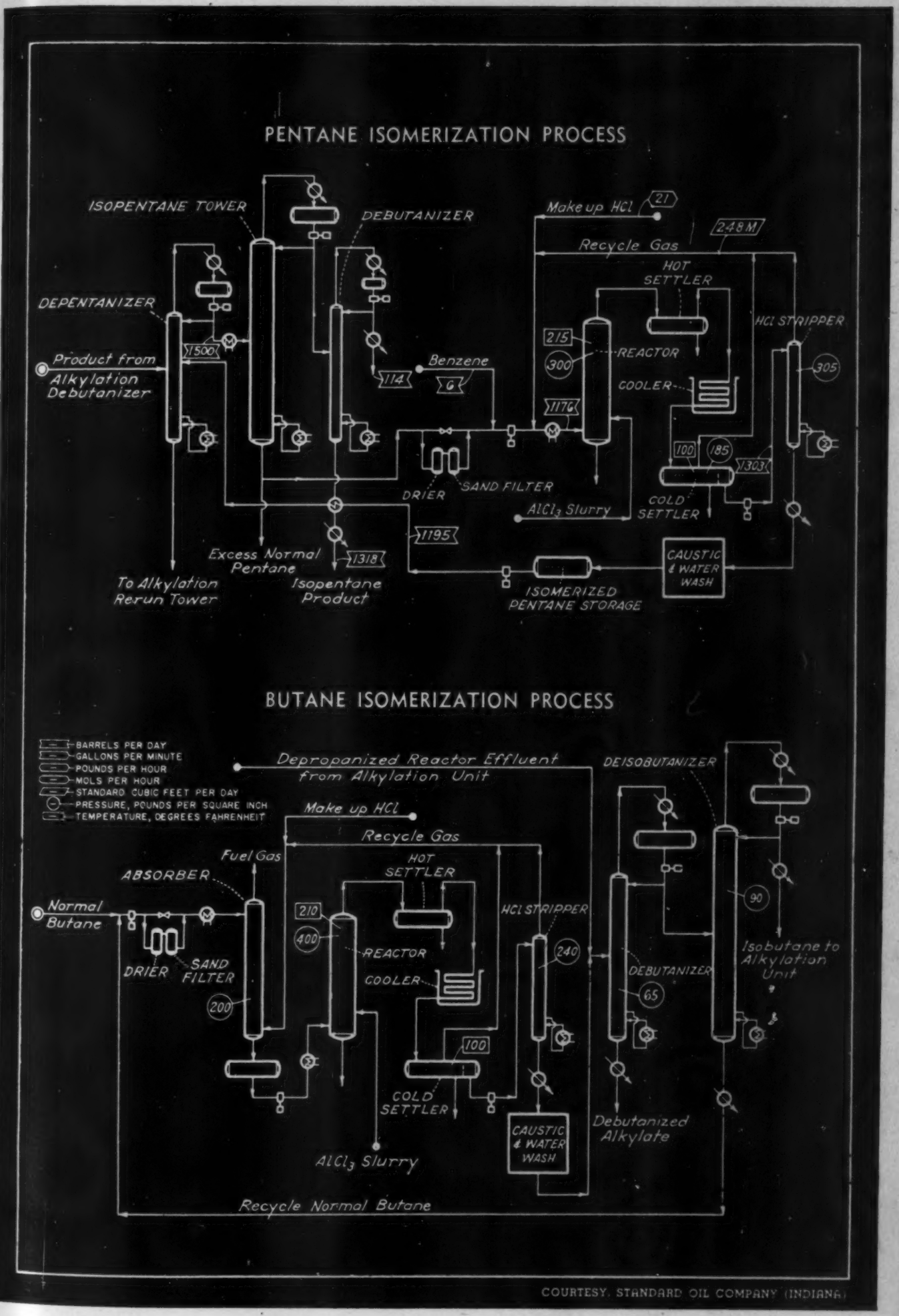
Hydrogen chloride from the hydrogen chloride stripper and a small amount of benzene are added to the normal pentane, and the solution is introduced through a distributor at the bottom of a column of aluminum chloride-hydrocarbon complex. The pentane isomate passing overhead from the reactor contains small amounts of entrained catalyst. A portion of this catalyst separates out in a hot settler. The effluent from the hot settler then is cooled, permitting further separation of catalyst in a cold settler. Hydrogen chloride not flashed off in the cold settler is stripped from the product in the hydrogen chloride stripper, and is recycled to the reactor. The isomerized pentane then is cooled and washed successively with caustic and water to remove the last traces of catalyst and hydrogen chloride.

Since the equilibrium is favorable to high once-through conversion, it is not always necessary to fractionate the product. However, if nearly pure isopentane is desired, the product can be passed to the fractionation system used for feed preparation. Thus, isopentane is removed overhead from the depentanizer, debutanized, and sent to isopentane storage.

A modification of this process can be used for the isomerization of normal butane to isobutane. Butane is dried and introduced into the top of a hydrogen chloride absorption system in a manner analogous to the Isomate Process. The normal butane solution of hydrogen chloride is passed upwardly through the aluminum chloride-hydrocarbon complex catalyst in the absence of added cracking inhibitors. The temperature and pressure used in the butane isomerization process are 210° F. and 400 pounds per square inch. The isomerized butane passes through settlers for the removal of traces of catalyst and thence to a hydrogen chloride stripper. The hydrogen chloride is removed overhead and recycled to the hydrogen chloride absorber. The stripped butane is caustic-washed and passed to a debutanizer, which may be that connected with the alkylation unit operating on the isobutane produced. The butane from the debutanizer passes to a butane splitter, the isobutane going overhead to an alkylation unit or storage, and the normal butane being recycled to the isomerization system.

Two commercial installations of this type have been constructed and operated successfully, one at the Whiting, Indiana, refinery and the other at the Wood River, Illinois, refinery of Standard Oil Company (Indiana).

Licenses for both the butane and pentane isomerization processes are available through Standard Oil Company (Indiana).



COURTESY, STANDARD OIL COMPANY (INDIANA)

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Ethyl benzene
Ethyl ether and isopropyl ether
Ethyl, isopropyl and other alcohols
Explosives and smokeless powder
Fats and fatty acids
Filtrol fractionated lubricating oils
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Nitroparaffines and derivatives
Penicillin
Petroleum products
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Phthalic anhydride
Propane deasphalted stocks
Pure hydrocarbons by fractionation or extraction
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Streptomycin
Synthetic acetic acid and acetic anhydride
Synthetic camphor
Synthetic phenol
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Toluene
Trinitrotoluene
Visbroken fuels

Somewhere along the line of Badger's experience can probably be found the aid your engineering department seeks to assure low manufacturing costs and dependable operation of the plant or unit you have in prospect.

Above is a partial list of chemical, petroleum and petro-chemical products in the commercial processing of which Badger offers wide experience.

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Liquid-Phase Isomerization Process

SHELL DEVELOPMENT COMPANY

THE Shell Liquid-Phase Isomerization Process is an entirely continuous regenerative process adaptable to the isomerization of either butane or pentane and commercial plants have proven completely satisfactory with both feeds. Some special merits of the process are: a constant high conversion for either hydrocarbon per pass coupled with a very low catalyst consumption, high ultimate yields of the desired isoparaffins, constant operating conditions, continuous catalyst make-up with continuous spent catalyst rejection, a high tolerance for feed impurities, and employment of standard fabricating methods for plant construction.

The process, in brief, consists of contacting n-butane or n-pentane in the liquid state with a catalyst comprising a solution of aluminum chloride in molten antimony trichloride. Dry hydrogen chloride gas is introduced into the feed hydrocarbon as catalyst promoter. In addition, when the process is utilized for pentane isomerization, hydrogen is introduced along with the hydrogen chloride to suppress disproportionation reactions which, if not controlled, would result in a low ultimate yield of isopentane and high catalyst consumption. The same plant can be employed for processing either of the aforementioned hydrocarbons by including a compressor to recycle the hydrogen required in the case of pentane isomerization. However, if a source of hydrogen at the desired pressure is available, such that it can be used on a once-through basis, even the compressor can be eliminated.

The simplified diagram on the opposite page shows the essential pieces of equipment and indicates the flow of the catalyst, hydrocarbon, activator and disproportionation inhibitor streams.

The normal liquid paraffin (butane or pentane) is charged continuously at essentially atmospheric temperature through one of two feed dryers, which are used alternately and renenerated periodically, through a heater where the temperature is raised to about 180° F., thence through a packed catalyst scrubber column and into the contactor. A liquid-liquid extraction of a small side-stream of the contactor catalyst with the feed takes place in the scrubber. The valuable active catalyst components pass back to the contactor in solution while the fluid inactive $AlCl_3$ hydrocarbon complex, which forms as a consequence of the small amount of side reactions, runs to the bottom of the column and is periodically drained off.

The contactor is a vertical autoclave equipped with a mixing device to provide intimate contacting of catalyst and hydrocarbon. A relatively short residence time, of the order of 10-15 minutes, in the contactor is sufficient to achieve a high conversion; 50 percent or better for butane and 55-60 percent for pentane. After phase separation of isomerizate from catalyst in the baffled upper settling section of the contactor, the isomerizate flows to the catalyst-removal column for separation of the dissolved catalyst by a simple fractionation. This recovered catalyst is pumped back to the contactor while the isomerizate is taken overhead, cooled, and passed to the accumulator from which it is pumped to the HCl stripper where the acid gas is removed and recycled to the contactor. The stripper bottoms are essentially free of HCl and after a caustic wash pass to subsequent operations.

In pentane isomerization the bulk of the hydrogen is recycled from the vapor phase of the catalyst column accumulator, however, some passes in solution to the HCl stripper and is recycled with the HCl.

The small amount of uncondensable gas that is formed is vented from the vapor phase of the catalyst column accumulator through a scrubber, countercurrent to a stream of HCl stripper bottoms which reabsorb HCl and reduce its loss to a minimum.

The constant high conversion, inherent to the process, is achieved by the continuous removal of spent $AlCl_3$ complex and regeneration of the contactor catalyst with fresh aluminum chloride. To accomplish the $AlCl_3$ addition a small portion of the bottoms from the catalyst removal column, which contain only a small amount of $AlCl_3$, is pumped through one of two $AlCl_3$ saturator vessels, filled with lump $AlCl_3$, before it is returned to the contactor. Since the $AlCl_3$ consumption in the process is quite low, this procedure serves to maintain the required concentration.

Since the catalyst is solid at atmospheric temperatures, lines carrying it are heat traced. Above its melting point, 165° F., the catalyst is quite fluid and easily transferred by pumping. All major pieces of equipment are of steel except those handling the catalyst which are nickel-clad. The temperatures of operation in the various units are moderate while the highest pressure necessary (HCl stripper) is around 325 psig.

This process is licensed by Shell Development Company, New York.

LIQUID PHASE PARAFFIN ISOMERIZATION PROCESS
1000 BARRELS PER DAY PRODUCTION OF ISOBUTANE OR ISOPENTANE
COURTESY SHELL DEVELOPMENT COMPANY
HCl MAKEUP
HCl RECYCLE
H_2 RECYCLE
H_2 MAKEUP
H_2 COMPRESSOR AND SURGE
VENT
VENT GAS SCRUBBER
HCl STRIPPER COLUMN
PRODUCT COOLER
CAUSTIC WASH
ISOMERIZATE OUT TO ALKYLATION OR ISOPENTANE SEPARATION
ISOBUTANE OR ISOPENTANE
ACCUMULATOR
HCl STRIPPER FEED PUMP
REFLUX
CATALYST REMOVAL COLUMN
CONTACTOR
CATALYST RECYCLE PUMPS
$AlCl_3$ SATURATORS
REJECT PUMP
COMPLEX OUT
CATALYST SCRUBBER COLUMN
FEED PREHEATER
DRIERS
FEED IN
FEED PUMP
CATALYST STORAGE AND UTILITY PUMP
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

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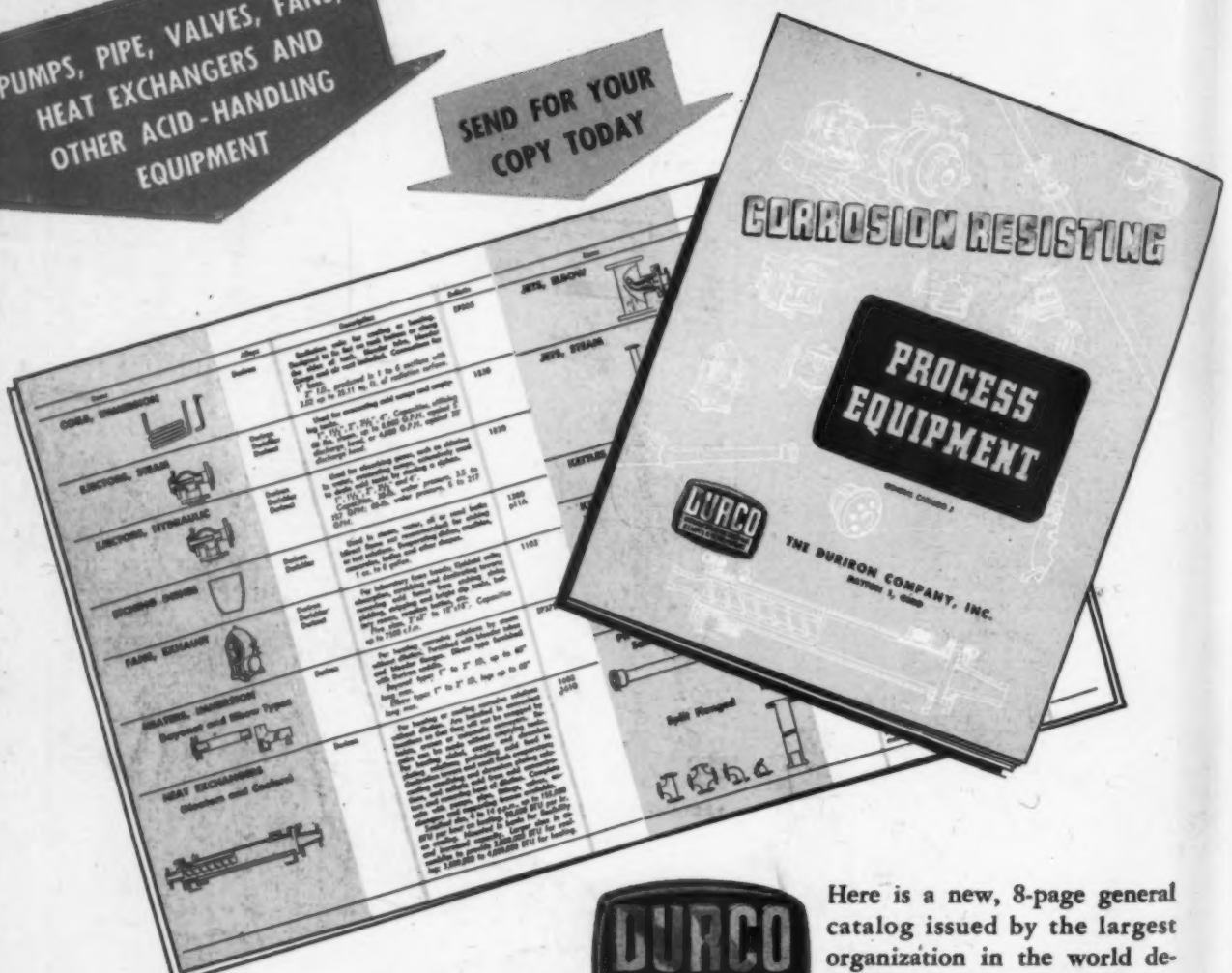

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Here is a new, 8-page general catalog issued by the largest organization in the world devoted exclusively to the production of *corrosion-resisting* alloys and equipment.

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Isomerization Process

UNIVERSAL OIL PRODUCTS COMPANY

THE UOP Isomerization Process was developed for the isomerization of normal paraffins to isoparaffins. This process has been used extensively for the isomerization of normal butane to isobutane which is one of the raw materials for making high octane gasoline blending component known as alkylate. The process is also applicable for the isomerization of normal pentane and hexane fractions for improvement of the octane number of these fractions.

The process involves the contacting of a dry feed with a catalyst which contains aluminum chloride and hydrogen chloride. The catalyst is deposited on the quartz chips packing in the reaction vessel and the activity of the catalyst is maintained at the desired level by the continuous introduction of fresh aluminum chloride dissolved in a portion of the liquid charge. Anhydrous hydrogen chloride is used as a promoter for the reaction and is recycled within the system to maintain the required concentration in the reaction zone.

The liquid hydrocarbon and chemicals composing the combined feed are contacted in their passage through the reaction zone. In the isomerization reaction a minor secondary reaction takes place, namely, the formation of a viscous liquid sludge containing dissolved aluminum chloride and hydrogen chloride which spreads over the packing surfaces. The sludge thus formed continuously drains from the packing and separates from the butane stream below the catalyst zone in the bottom section of the reactor, and this small amount of material is drawn off intermittently to a neutralizing pit. The hydrocarbon effluent from the reaction zone, containing dissolved aluminum chloride and hydrogen chloride, then enters the aluminum chloride and hydrogen chloride recovery section.

The isomerization reaction is slightly exothermic. The heat generated in the process is absorbed by the vaporization of some of the liquid passing through the reactor, and as a result the reaction is essentially isothermal. The process is operated at pressures in the order of 250-350 psi and temperatures of 190° F. to 220° F.

The effluent from the reactor is fed to an aluminum chloride separation column where the net hydrocarbon product, together with recycled hydrogen chloride is separated from the aluminum chloride which is concentrated in the bottoms stream from the column, consisting mainly of pentanes and normal butane.

The hot stream from the bottom of the aluminum chloride column containing between 1 and 3 pounds of aluminum chloride per barrel is returned to the top of the reaction vessel. Recycling aluminum chloride in this manner makes it possible to maintain a high concentration of aluminum chloride in the reactor which could not otherwise be possible without excessive chemical consumption. A small portion of the bottoms stream from the aluminum chloride column is withdrawn to a caustic wash tower to reject continuously the small amount of pentanes formed in the process, which otherwise would build up in the system.

The overhead from the aluminum chloride column is fed to a conventional HCl stripper to separate the HCl from the net plant product. The HCl from the top of the HCl stripper is returned to the reactor as a recycle stream. The bottoms stream from the HCl stripper is caustic washed to remove any traces of hydrogen chloride before leaving the unit.

The net product from the plant consisting of a mixture of isobutane and normal butane is sent to a deisobutanizer where the unconverted normal butane is separated and returned to the isomerization unit. Since the isomerization unit normally operates in conjunction with an alkylation unit, the isobutane-normal butane separation generally is carried out in the fractionation train of the alkylation unit which separates an isobutane rich stream which in turn is fed to the alkylation reactors.

The isomerization of pentanes or hexanes is carried out in substantially the same manner as that described for the processing of normal butane with the exception that in the processing of these heavier materials there is a greater tendency for the formation of sludge, and this must be counteracted by the injection of an inhibitor which serves to suppress secondary sludge-forming reactions.

Typical Isomerization Results
Normal Butane, Pentane and Hexane Feeds

TYPE OF FEED	n-Butane	n-Pentane	Hexane Fraction
Operating Conditions:			
Reactor Temperature, °F.	190	180	180
Reactor Pressure, psig.	250	250	250
HCl in Reactor Feed Mol., Percent	7-10	7-10	7-10
Product Analyses:			
Isobutane	50	....	....
Normal Butane	48	....	....
Isopentane		60	
n-Pentane	2	38	
Hexane and Heavier		2	
Octane Ratings:			
Feed ASTM O.N. Clear			58-60
Product ASTM O.N. Clear			75-80
3 cc TEL/gal.			88-93
Catalyst Consumption:			
Lbs. $AlCl_3$/bbl. iC_4 or iC_5 produced	0.3-0.5	0.5-0.7	
Lbs. HCl/bbl. iC_4 or iC_5 produced	0.05	0.05	

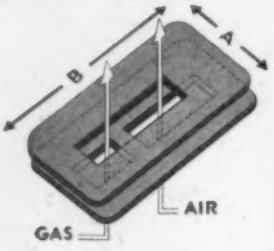
NORMAL BUTANE, PENTANE OR HEXANE FRESH FEED
RECYCLE
REACTOR
HCL RECYCLE
ALUMINUM CHLORIDE SATURATOR
ALUMINUM CHLORIDE TOWER
SLUDGE
HCL COLUMN
ISOMERIZED PRODUCT
CAUSTIC
SPENT CAUSTIC
U.O.P. ISOMERIZATION PROCESS
COURTESY, UNIVERSAL OIL PRODUCTS COMPANY

Kemp Inert Gas Generators

HUNDREDS of millions of dollars worth of equipment and products in both producing and refining fields are protected from fire and explosion by inert gas produced in Kemp Inert Gas Producers.

The self-contained, low maintenance, automatic Kemp Inert Gas Generators provide 1,000 to 120,000 cfh of Inert Gas which can be stored under pressure or piped to many points throughout the plant, or merely made as needed. In any event, an adequate supply of gas is available at all times without danger of shortage due to errors, delivery problems, or other adventitious causes.

Kemp Generators make inert gas by the complete combustion of gaseous fuels and controlled processing of the products of combustion. The composition of the inerts depends on the fuel used and the ratio of air with which it is mixed. 12½% CO_2, 87% N_2 and ½ of 1% residuals, either oxidizing or reducing as desired, is a typical product.

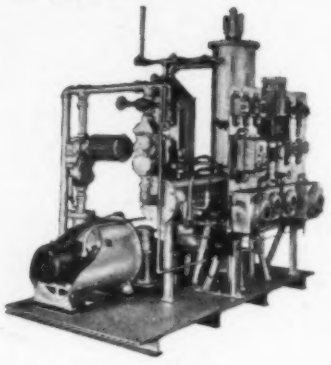


The Kemp carburetor is fully automatic and works on the slide-valve principle illustrated in the small diagram at left. The slides move in the direction B-B, controlled by lockable vernier micrometer handwheel adjustment, to set the mixture. They move in the direction A-A, automatically controlled by demand acting through diaphragm and yoke lever, to increase or decrease the delivery of the fuel-air mixture.

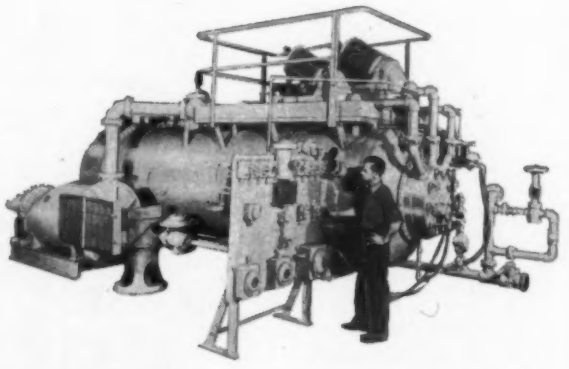
Average costs per Mcf are 10¢ for unprocessed inerts, 12¢ for compressed inerts, 17¢ for desiccated inerts. Low pressure nitrogen averages 16¢, compressed nitrogen 18¢, and desiccated nitrogen 23¢ per Mcf. These figures are based on a medium-size generator, low pressure delivery at 15" water column, high pressure delivery at 100 psi, and cooling to within 10° F. of cooling water temperature. Bottled gases, setting aside the expense and nuisance of handling, labor, cylinder inventory, etc., frequently cost 100 times as much as gases made in your own plant with Kemp equipment. Moreover, there is no waste in Kemp-generated inerts. Fuel and power costs are exactly in proportion to demand. You make and pay for exactly what you need. Kemp Inert Gas Generators are virtually automatic in operation, virtually free of maintenance for many years, and provided with every necessary protection against fire hazard. They are simple and rugged in design, fully accessible to anyone with a screwdriver and a wrench, and operate at full efficiency without the services of an engineer.

Kemp Safety Accessories protect against every hazard of flash-back in every kind of piping-system. Kemp flame-arresters, firechecks (manual and automatic), backfire preventers (indoor and outdoor types), explosion softheads, and safety-rod vents are widely used throughout industry and carry Underwriters' Laboratories label. Kemp engineers will gladly advise on installation.

Send for complete information on Kemp Inert Gas Generators in Bulletin I-10.



3 MID Kemp Inert Gas Generator used by major western oil producer for line purging.



50 MIHE Kemp Inert Gas Generator supplying purging and flushing requirements at a midwestern oil refinery.

Specifications Kemp "HE" Inert-Gas Generators

			CONNECTIONS			DIMENSIONS			
MODEL	CAP. C.F.H.	MOTOR HP	FUEL	COOLING WATER	INERT-GAS DISCHARGE	L	W	HT.	SHIPPING WEIGHT
1 MIHE	1,000	1	1 "	1 "	1½"	6' 6"	3' 6"	7' 0"	3,000 LBS.
2 MIHE	2,000	2	1 "	1 "	2 "	6' 6"	4' 0"	7' 0"	3,500 "
3 MIHE	3,000	2	1 "	1½"	3 "	8' 0"	4' 6"	7' 6"	4,500 "
4 MIHE	4,000	2	1 "	1½"	3 "	8' 0"	4' 6"	7' 6"	4,500 "
6 MIHE	6,000	3	2 "	1½"	3 "	11' 6"	6' 0"	8' 0"	6,000 "
8 MIHE	8,000	3	2 "	2 "	4 "	11' 6"	6' 0"	8' 0"	10,000 "
10 MIHE	10,000	5	2½"	2 "	6 "	14' 0"	7' 6"	10' 0"	12,000 "
15 MIHE	15,000	5	2½"	2 "	6 "	14' 6"	7' 6"	10' 0"	16,000 "
20 MIHE	20,000	10	3 "	3 "	6 "	14' 6"	7' 6"	10' 0"	18,000 "
30 MIHE	30,000	10	4 "	3 "	8 "	17' 0"	9' 0"	10' 6"	23,000 "
50 MIHE	50,000	15	4 "	4 "	10 "	20' 0"	10' 0"	11' 0"	30,000 "
100 MIHE	100,000	30	4"-6"	4 "	12 "	26' 0"	15' 0"	12' 0"	45,000 "

Specifications Kemp "D" Inert-Gas Generators

			CONNECTIONS			DIMENSIONS			
MODEL	CAP. C.F.H.	MOTOR HP	FUEL	COOLING WATER	INERT-GAS DISCHARGE	L	W	HT.	SHIPPING WEIGHT
1 MID	1,000	1	1 "	1 "	1½"	6' 0"	5' 0"	9' 0"	2,500 LBS.
2 MID	2,000	2	1 "	1 "	2 "	7' 0"	5' 9"	9' 6"	3,000 "
3 MID	3,000	2	1 "	1½"	3 "	7' 0"	5' 9"	9' 6"	3,500 "
4 MID	4,000	2	1 "	1½"	3 "	7' 0"	5' 9"	9' 6"	3,500 "
6 MID	6,000	3	2 "	1½"	3 "	13' 0"	6' 0"	15' 0"	6,000 "
8 MID	8,000	3	2 "	2 "	4 "	13' 0"	6' 0"	15' 0"	10,000 "
10 MID	10,000	5	2½"	2 "	6 "	16' 0"	7' 0"	15' 0"	12,000 "
15 MID	15,000	5	2½"	2 "	6 "	17' 0"	7' 0"	16' 0"	16,000 "
20 MID	20,000	10	3 "	3 "	6 "	17' 0"	7' 0"	16' 0"	18,000 "
30 MID	30,000	10	4 "	3 "	8 "	20' 0"	8' 0"	16' 0"	23,000 "
50 MID	50,000	15	4 "	4 "	10 "	24' 0"	8' 0"	17' 0"	30,000 "

HE-type generators cool the inert gas in a heat exchanger. They are preferred for critical chemical process work or whenever complete freedom from oxygen is imperative. In D-type models the gases are actively scrubbed by direct contact with water sprays; often preferred when sulphurous gases must be removed from inerts made of sour fuel.

Kemp "Dynamic" Dryers

ONE authority says "Corrosion is a one-hundred-million dollar per year problem in the petroleum industry." Kemp "Dynamic" Dryers are licking that problem in many petroleum refining and producing operations. Chief uses are: (1) humidity control for equipment and spaces to prevent mold, rust, corrosion, moisture absorption by product; (2) drying of gases (including compressed air) to sub-zero dew points to prevent condensation and clogging of instruments; and (3) desiccation of non-aqueous liquids.

Kemp Dryers are available as "standard" or "built-to-order" units. Yet "built-to-order" equipment, however complex and exacting the requirements, incorporate so much of Kemp "standard" design as to assure investment economies to the purchaser.

Capacities range from 1 to 20,000 cfm; operating pressures from atmospheric to 2500 psi. Final dew points as low as —60° F. may be reached. Operation is manual, semi-automatic, or fully automatic. Gas, steam, or electricity may be used to supply the heat for reactivation of the adsorbent.

The typical cycle is one of adsorption, reactivation, and cooling. Wet gas enters the drying tower and passes through a bed of silica-gel or activated alumina, the depth of which is determined by the degree of dryness which is sought. After the adsorbent tower has exhausted its designed capacity (determined by time on the line) the gas flow is directed to the twin tower. Ambient air is now driven through a heater and through the first tower at 300° to 350° F.; thence through a cooler and to the atmosphere. When the bottom of the desiccant bed reaches reactivation temperature all water has been removed. Air inlet and outlet are now closed and the heater shut off so that one volume of air is continuously recirculated through tower and cooler until the desiccant is restored to room temperature and ready to go back on the line when needed. Usual cycles run from 8 to 12 hours.

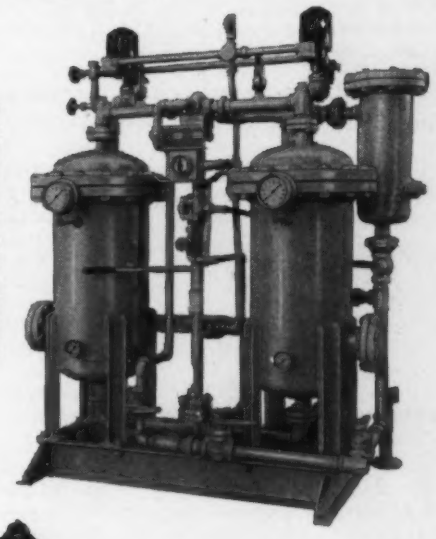
Costs are low! Compressed air may be dried to sub-zero dew points for ¼ to ⅓ of 1¢ per Mcf when capacity of over 3,000 cfm is available. In the 300 to 2,000 cfm range cost is 1¢ to 2¢ per Mcf. A top figure for small units without pre-cooling or refrigeration is 4¢ to 6¢ per Mcf.

Kemp "Dynamic" Dryers are adapted to handle hydrocarbons, refrigerants, organic intermediates, solvents, aromatics and other liquids.

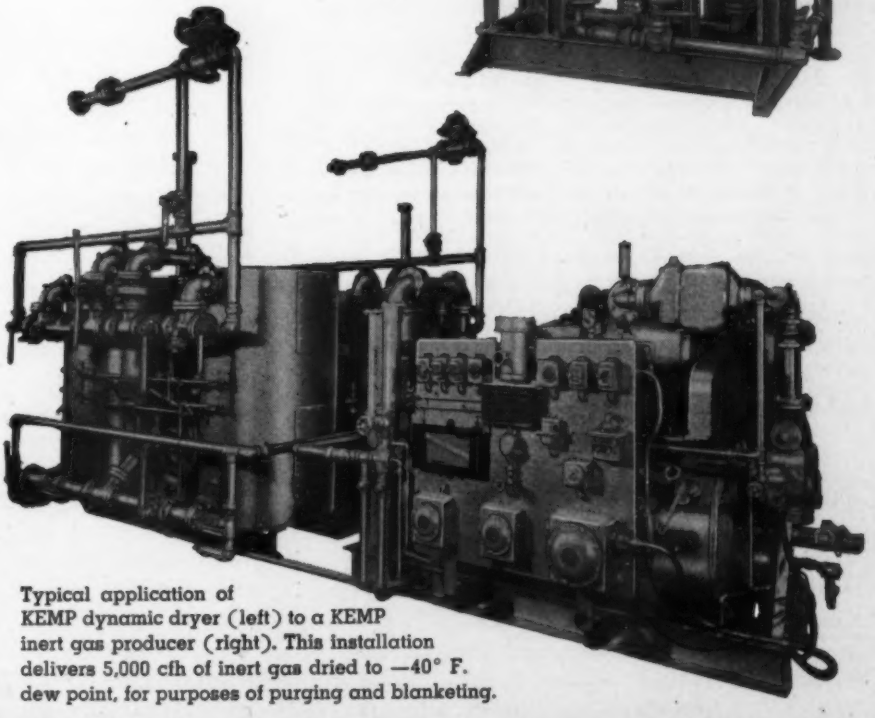
The operation of Kemp Dryers is extremely simple. No engineering skill is required; any intelligent workman can be instructed in a day.

Installation, too, is relatively simple though a Kemp technician is always at hand to make the basic settings and adjustments.

Design or selection of equipment is the foundation of full success in service. Here is where Kemp engineers give weight to many factors: quantities and materials to be dried, speed and degree of drying, diameter and depth of drying beds, pressure, and many other variable factors. Experience alone—and we have years of it—guides the weighting of these factors in achieving profitable trouble-free operation. That experience is at your service. Send for a Kemp engineer or ask to see Bulletin 25-D.



Typical Kemp compressed air drying unit of the ORIAD (oil refinery instrument air dryer) type designed for 100 cfm of air at 80 psi and 70° F. dried from saturation to a dew point of —30° F.



Typical application of KEMP dynamic dryer (left) to a KEMP inert gas producer (right). This installation delivers 5,000 cfh of inert gas dried to —40° F. dew point, for purposes of purging and blanketing.

The C. M. KEMP Mfg. Company

405 EAST OLIVER STREET, BALTIMORE 2, Md.

Isomate Process

STANDARD OIL COMPANY (INDIANA)

THE Indiana Isomate Process is a method for converting low-octane-number pentanes and hexanes into isomers of higher octane number. Liquid aluminum chloride-hydrocarbon complex promoted with anhydrous hydrogen chloride is used as a catalyst. Hydrogen is added to repress cracking and disproportionation and to maintain high catalyst activity. Conversion is accomplished by passing the liquid hydrocarbon charge in a dispersed phase upwardly through the catalyst complex. The preferred operating conditions are 240-250° F. and 700-800 pounds per square inch. Suitable charging stocks for the Isomate Process are light, straight-run naphthas containing pentanes and hexanes and but little heavier hydrocarbons.

In addition to handling mixtures of pentanes and hexanes, the process is applicable for the isomerization either of pentanes alone or hexanes alone if the refinery situation indicates such operations to be desirable. However, for the isomerization of pentanes alone, the Indiana Pentane Isomerization Process is somewhat more economical.

When charging a mixture of pentanes and hexanes, the Isomate Process can be operated on a once-through basis to manufacture a product of 80 unleaded CFR-M octane number. Exceptionally high liquid yields, in the order of 100 volume-percent on charge, can be obtained in this type of operation. By fractionating the once-through isomate and recycling the low-octane-number pentanes and hexanes to the reaction zone, the quality of the product can be raised to 91 unleaded CFR-M octane number with but a slight sacrifice in yield. When recycling is employed with a charge stock consisting primarily of paraffinic hexanes, the principal product is neohexane.

Isomate plants of approximately 5000 barrels and 1750 barrels per day capacity have been constructed and successfully operated at Whiting, Indiana, and Salt Lake City, Utah. Pentane recycling was employed at Whiting and hexane recycling at Salt Lake City.

Referring to the flow diagram, naphtha feed passes first through a calcium chloride dryer or other drying means, a sand filter, and thence to the top of a hydrogen chloride absorption tower. In the absorption tower hydrogen chloride is recovered from the recycle gas. A small amount of makeup hydrogen chloride is added as required at this point. The naphtha containing dissolved hydrogen chloride passes from the accumulator at the bottom of the absorption tower through a preheater to the reactor. The naphtha and gaseous hydrogen are introduced simultaneously through a distributor at the bottom of the reactor, which is an open vessel filled with aluminum chloride-hydrocarbon complex catalyst.

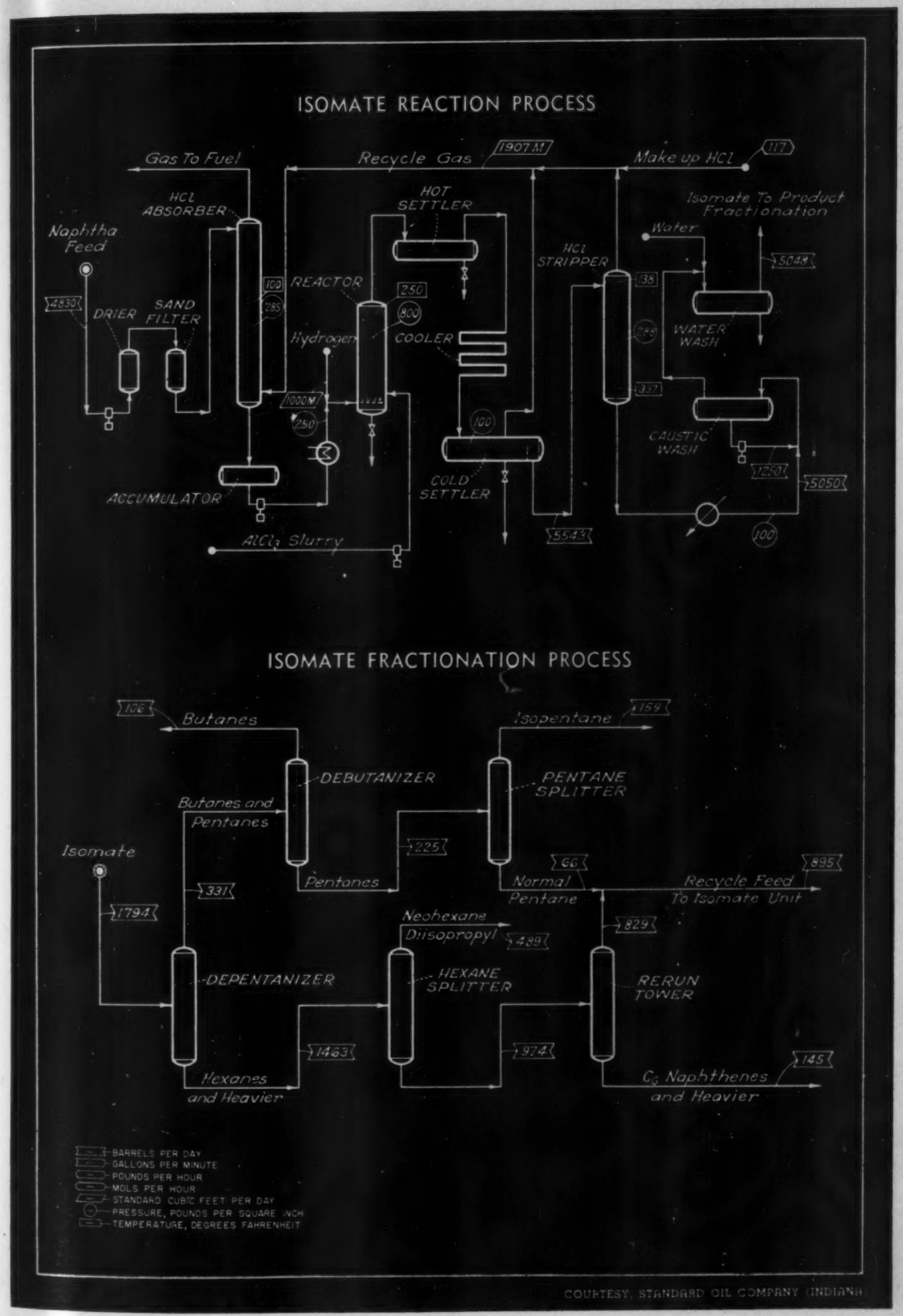
The hydrocarbons, being considerably lighter and substantially immiscible with the catalyst, pass upward through the catalyst. The effluent from the reactor, consisting of isomerized naphtha, hydrogen chloride, hydrogen, and a small amount of entrained catalyst, passes successively through a hot settler, a cooler and a cold settler. The catalyst is separated in the two settlers and returned to the reactor or to catalyst storage; hydrogen and hydrogen chloride are flashed off in the cold settler, and hydrogen chloride remaining in the effluent from the cold settler is removed in the hydrogen chloride stripper. The liquid product from the bottom of the hydrogen chloride stripper passes through a cooler and then successively through caustic and water washes to eliminate the last traces of catalyst and hydrogen chloride. Hydrogen chloride from the stripper is recycled to the absorber. The isomate product can be debutanized for blending into aviation fuel or blended directly into motor fuel.

In some instances, it may be desirable to further fractionate the isomate. The isomate fractionation system used in connection with the smaller installation is also shown on the flow diagram.

Crude isomate is passed first to a depentanizer in which butanes and pentanes are taken overhead and hexanes, along with traces of heavier materials, are removed as bottoms. The overhead butanes and pentanes are separated by fractionation, the pentanes passing to a pentane fractionator from which substantially pure isopentane is taken overhead, with normal pentane eliminated as bottoms for possible recycle to the isomate unit. The hexanes and heavier materials from the bottom of the depentanizer pass to a hexane fractionator where neohexane and diisopropyl are removed as overhead products; the bottoms are sent to a rerun tower where cyclohexane, and heavier materials unsuitable as charging stocks for the isomate process are removed. The overhead methylpentanes and normal hexane from the rerunning operation may be combined with normal pentane and recycled to the isomate unit.

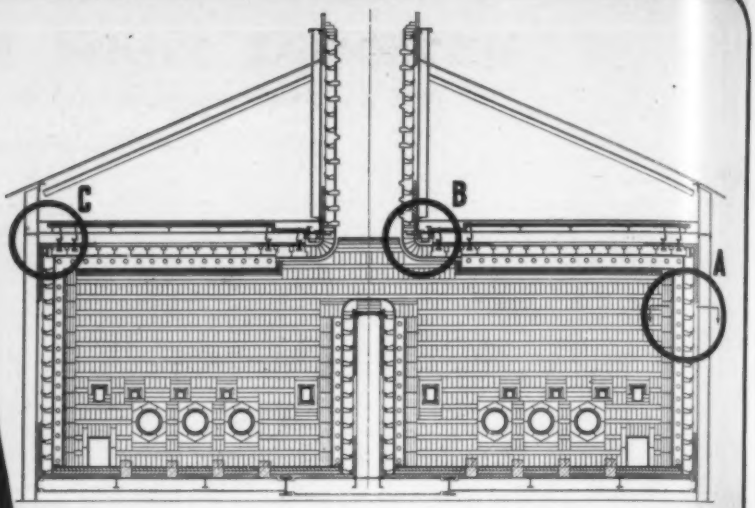
Products from the Isomate Process are primarily of interest as blending stocks for high-octane aviation fuels. However, the process is very suitable for improving the octane number of light straight-run naphthas for motor fuels where low-octane-number charging stocks such as normal pentane and normal hexane are available, or where octane ratings must be increased beyond levels attainable with existing equipment.

Licenses for operation of this process are available from Standard Oil Company (Indiana).

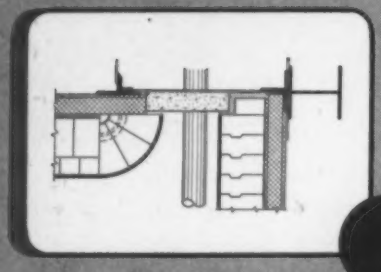


COURTESY, STANDARD OIL COMPANY (INDIANA)

WHERE BIGELOW-LIPTAK ENGINEERING answered 3 critical construction problems

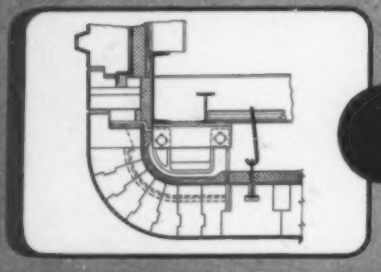


The installation pictured above is one of two identical delayed coker furnaces which have been in operation on the west coast for two years. Specifications called for Bigelow-Liptak engineering utilizing unit-suspended wall and arch construction with special emphasis at three potential "trouble spots."



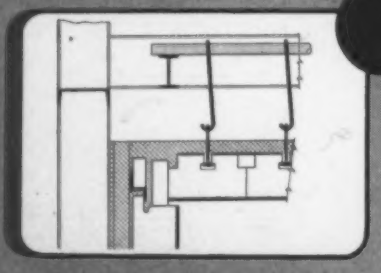
A

Offset construction, as shown at the left, provides a positive seal despite the differences in expansion movements of the unit-suspended wall and the tube sheets. Radial tile replace the vulnerable sharp corners. A minimum of tube sheet protective castable is exposed and its corners are protected.



B

Substitution of a *free floating* nose for a rigidly hung assembly eliminates tile and casting failures. Self-aligning castings permit the entire nose to accommodate expansion movements without stress.



C

The free swinging or pendulum type of arch support is carried through the entire flat arch giving the same advantages as the nose described above. Offset construction at these corners stops leakage and shields expansion joint material from hot furnace gases.

There are many types of standard Bigelow-Liptak construction. This was a no-casing job, using the 9" double wall, the standard 7" offset wall, and the 7" offset arch. More information on industrial furnace enclosures is available upon request.

write to... **BIGELOW-LIPTAK CORPORATION**

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By means of pickup tube, AER-O-FOAM Liquid may be used direct from the original 5-gallon container (as illustrated). Other means of producing AER-O-FOAM are available for portable or fixed system service.

AER-O-FOAM is a mechanical foam, produced from a proper mixture of AER-O-FOAM Liquid, water and air by equipment especially designed and developed by National Foam System, Inc. AER-O-FOAM Liquid may be used with either salt or fresh water in any unit.

National Foam Powders are packed in hermetically sealed steel pails (as illustrated) with quick opening ring seals and closed under 800 lbs. pressure to prevent deterioration. For easy identification, the background of the "A" pail is red, the "B" pail black, and "2 in 1" yellow.

National Foam "99" Foam Powder (as illustrated) fully protects alcohol, acetone, ether and other organic solvents. It is equally effective with alcohol-gasoline blends.

We have had years of experience in the development, design and production of foam producing chemicals and mechanical devices for economical and efficient application of foam.

Write for catalogue and let us suggest equipment best suited for your needs. The services of our Engineering Department are available without obligation.

Isoforming Process

STANDARD OIL COMPANY (INDIANA)

THE Indiana Isoforming Process is suitable for the improvement of thermally-cracked naphthas, especially those of high olefin content. The hydrocarbon vapors are passed over solid catalysts at high space velocities. Although the improvement in octane rating is moderate, extremely high yields are obtained.

The degree of improvement in octane number in the Isoforming Process depends primarily upon the olefin content of the stock and the octane number level of the olefins. A high olefin content and low octane number are conducive to greater octane number improvements. A typical improvement for a Mid-Continent thermally cracked naphtha is 4 to 5 units, for example, 68 to 72 ASTM clear. Coke-still naphthas which have a higher olefin content and lower octane number can be improved 8 to 10 units, for example, from 59 to 68 ASTM clear. The process yields about 97.5 to 98.5 volume-percent liquid product based on charge. Small losses to dry gas and carbonaceous residues amount to about 1.8 percent based on volume charged. About 2.5 percent of a polymer boiling above the end point of the charging stock is formed. In addition to octane number improvement, there is an appreciable desulfurization, which improves the lead response.

The process is carried out at relatively low pressures, for example, 10 to 20 pounds per square inch, at temperatures of about 950° F. Space velocities of from 75 to 125 barrels per hour per ton of catalyst can be used. In some instances, even higher space velocities are practicable. The process makes use of solid catalysts and fixed-bed operation. The activity of the catalyst is maintained for a period of about 48 hours, after which the catalyst requires regeneration. It is regenerated by burning off carbonaceous deposits with a flue gas containing an average of 5 mol-percent of oxygen. The maximum regeneration temperature is 1050° F. In order that the process be continuous, it is necessary to have two reaction chambers operating in parallel, one being regenerated, while the other is on stream.

Several types of catalyst are suitable. Acid-treated clays can be used as can synthetic catalysts of the silica-alumina type. Synthetic zeolites are especially effective and are preferred.

The flow diagram on the opposite page illustrates the operation. The heavy naphtha charge is heated first by heat exchange and then by a furnace to the desired temperature. It passes as a vapor into the reactor that is on stream. The "isoformate" and small traces of polymer and gas pass from the reactor to the fractionator. Polymer is removed as fractionator-bottom product; the overhead consisting of dry gas and "isoformate" passes to a separator, the gas being vented to an absorption plant and the liquid condensate passing to a depropanizer. The "isoformate" product is removed as bottoms from the depropanizer.

A gas generator operating on air and fuel gas provides a flue gas containing about 5 percent oxygen, which is used for regeneration in the other reaction chamber. The effluent gas stream from the regeneration passes to a waste-heat boiler for heat recovery, and thence to the stack.

Isoforming units have been designed to process 20,000 barrels per stream day of heavy cracked naphtha. No commercial installations have been made, since commercialization was interrupted by the war. It is only now that reconsideration is being given to their installation. The Isoforming Process should perform an important function in small refineries which have not yet installed catalytic cracking, in addition to being useful in all refineries where large volumes of coke-still naphthas and other thermally-cracked stocks are available.

Licenses for this process are available from Standard Oil Company (Indiana).

ISOFORMING PROCESS
Heavy Naphtha Charge
20000
Gas To Absorption Plant
DEPROPANIZER
1900
Isoformate
500
Polymer
FRACTIONATOR
1000
REACTOR
On Stream
350
15
REACTOR
Regenerating
WASTE HEAT BOILER
Flue Gas
Vent
Fuel
Air
GAS GENERATOR
BLOWER
FURNACE
Superheated Steam
Water
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
COURTESY, STANDARD OIL COMPANY (INDIANA)

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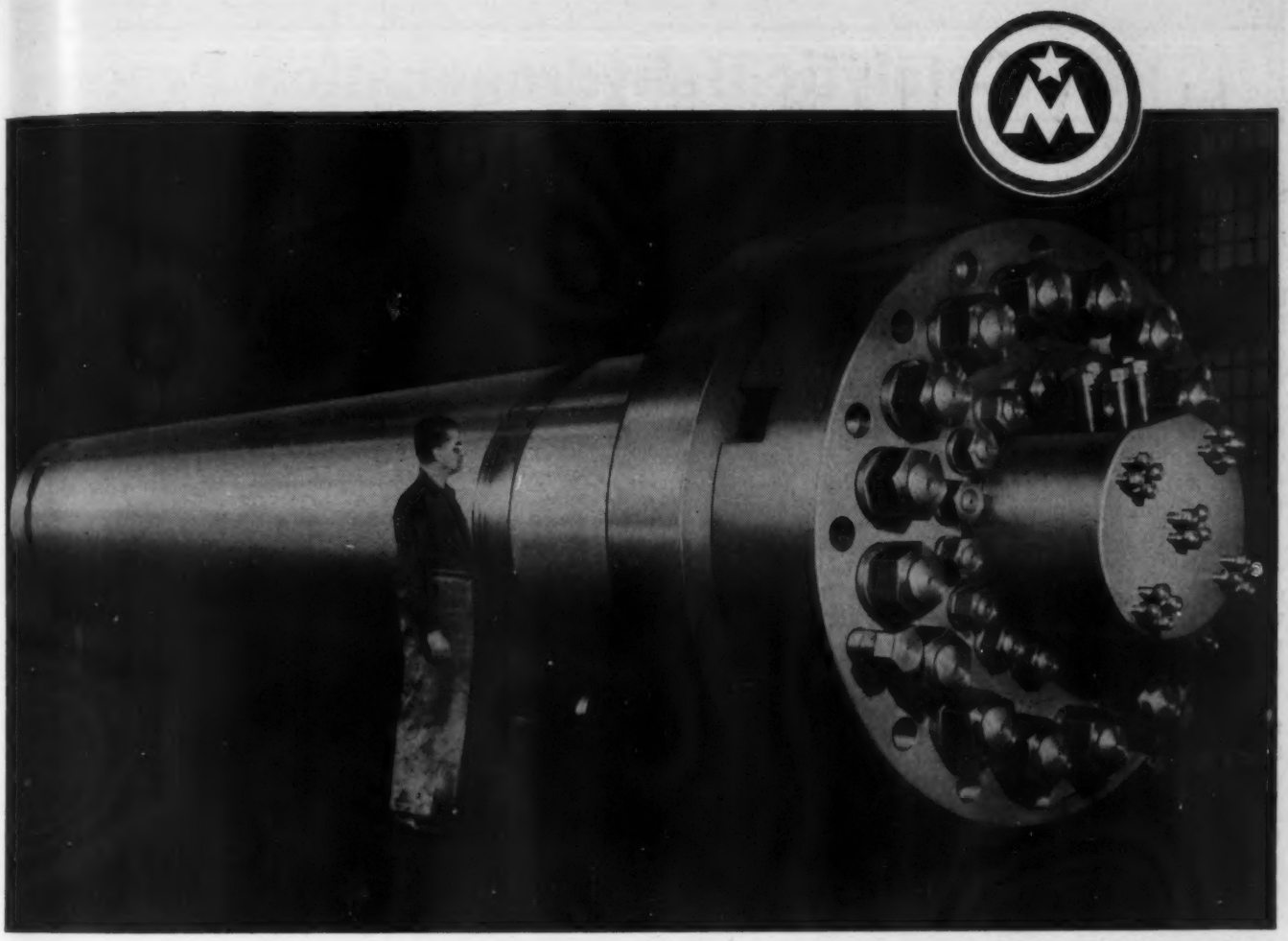
Howard Harmon	Howard E. Gibson	F. M. Tyler & Co.	General Equipment Co.	H. L. Lowry



The chrome vanadium steel ingot from which this Pressure Vessel was forged weighed approximately 550,000 pounds.

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Midvale makes one piece hollow Forged Pressure Vessels for the chemical, refining and process industries under the watchful eye of the Midvale quality control system. Midvale quality control starts with the design and is maintained through all operations to the finished product.

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Custom Steel Makers to Industry

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•

CORROSION AND HEAT RESISTING CASTINGS

•

FORGINGS AND RINGS

Catalytic Dehydrogenation Process

UNIVERSAL OIL PRODUCTS COMPANY

THE UOP Catalytic Dehydrogenation Process, broadly speaking, covers the removal of hydrogen from hydrocarbons to produce less saturated hydrocarbons. Such hydrocarbons, depending upon the degree of unsaturation, are useful in the production of fuels for internal combustion engines and as basic materials in the synthesis of rubber, resins and various chemicals. Catalytic butane dehydrogenation means removal of hydrogen from butane to form butylene and hydrogen.

The catalytic butane dehydrogenation process produces large yields of butylene due to the high efficiency of the catalytic reaction. The catalyst speeds up the desired primary reaction of dehydrogenation without a corresponding increase in the reactions which produce undesired secondary products.

The reaction is carried out in reactor tubes containing a suitable catalyst at temperatures ranging from approximately 1000° F. to 1100° F. and pressures ranging from approximately 10 to 50 psi.

The dry normal butane charge is heated to reaction temperature and introduced into the catalyst zone where it undergoes dehydrogenation, drawing the required heat of reaction from an external circulating flue gas system.

The material leaving the catalyst tubes is composed of unreacted butane, butylene, hydrogen, and a small percentage of other light gases such as methane, ethane, ethylene, propane, and propylene. It passes through the butane charge heat exchanger into a gas scrubber where it is cooled by a circulating oil stream and washed free of any entrained tar or coke. The gas then is compressed to 100-150 psi and discharged to the gas recovery system for separation of the butane-butylene fraction from the hydrogen and light gases.

As the dehydrogenation reaction proceeds, a small amount of carbon is produced which deposits on the catalyst and reduces its activity. This carbon must be periodically burned off by circulating oxygen-containing flue gas over the catalyst.

In order that the processing may be continuous, the reactors containing the catalyst are built in pairs, and by means of suitable, automatically-controlled switch valves, one reactor is kept in service dehydrogenating butane, while the other is being reactivated by having the carbon burned off.

The reactivating system is similar to the processing system and runs parallel with it. A reactivating gas of controlled oxygen content and sufficient volume is used to permit burning the carbon from the catalyst without reaching excessively high temperatures.

The reactivating gas leaves the compressor and passes through a heat exchanger and heater; it is then directed by means of switch valves to the catalyst tubes in the spent reactor. The inlet temperature is automatically controlled by firing valves on the reactivating gas heater and is normally carried at 800° F.

The burning of the carbon from the catalyst produces heat which must be dissipated. Part of this heat is removed by raising the temperature of the reactivating gas itself, the remainder by the external flue gas circulating system.

The reactivating gas leaving the catalyst tubes passes through a heat exchanger and into a gas washer, where it is cooled by water and then returned to the compressor for recirculation. Excess gas, corresponding to the air added, must be discharged from the system in order to prevent a pressure rise on the system. This gas is released by means of a pressure control valve which maintains constant pressure on the system.

The oxygen content of the reactivating gas is controlled between 1 and 3 percent to avoid excessive temperatures during the carbon-burning period. This is accomplished by means of an oxygen controller, which takes a continuous sample of the circulating reactivating gas stream and determines the oxygen content. As this varies, the controller operates the inlet air valve to admit the amount of air necessary to maintain the oxygen content at the desired figure.

The circulating flue gas system, which serves to supply heat to the processing reactors and remove heat from the reactivating reactors, is composed of a blower, a heater, a stack, switch dampers and interconnecting ductwork. Part of the circulating flue gas passes from the blower through the heater, where it is mixed with hot combustion gases to bring it to the desired temperature, after which it passes in indirect heat exchange with the processing reactor. The other part of the circulating flue gas passes through the cooling duct, where it is either sprayed with water or mixed with cool air; it then passes in indirect heat exchange with the reactivating reactor. After passing through the reactors, the heating and cooling flue gases combine in a common outlet duct and return to the blower. Pressure on the circulating flue gas system is controlled by discharging excess gas to the stack.

The commercial dehydrogenation of normal butane is usually carried out at a conversion of about 30 percent per pass in a recycle operation. The ultimate yield of butylene is 75 to 80 barrels per 100 barrels of butane converted.

In practice, the butane-butylene fraction usually has been fed to an alkylation or a polymerization unit in which the butylenes are converted to alkylate or polymer for the production of high octane gasoline, while the unreacted butane is recovered for recycling to the dehydrogenation unit.

CATALYTIC DEHYDROGENATION PROCESS
SWITCH VALVES
REACTIVATING GAS HEATER
R. G. WASHER
TO VENT
R. G. COMPRESSOR
R. G. EX CHANGER TO DRAIN
WASHER WATER
AIR
PROCESS GAS SCRUBBER
SCRUBBER OIL COOLER
PROCESS GAS COMPRESSOR
SCRUBBER OIL PUMP
GAS TO ABSORBER
RECEIVER
PRECOOLER
LIQUID TO STRIPPER
ABSORBER BOTTOMS
BUTANE HEATER
BUTANE HEAT EXCHANGER
GUARD TUBE
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Petrol Bronze
Petrol Bronze K
Petrol Bronze ST
Petrol Nugget Bronze
Petrol Amber R
Petrol Totem Bronze
Petrol Vermillion
Petrol Bronze R7 *(Reddest)*

GREENS

Petrol Traffic Green *(Bluest)*
Petrol Green N. F.
Petrol Green 2G
Petrol Green N. F. G.
Petrol Green D
Petrol Green 3Z
Petrol Green ST
Petrol Green 3W
Petrol Green CSN
Petrol Green MR
Petrol Green N. S. G.
Petrol Green CS
Petrol Green K2 *(Yellowest)*

PURPLES

Petrol Purple No. 4 *(Reddest)*
Petrol Purple 3R
Petrol Purple SC
Petrol Kero Purple N. A.
Petrol Purple No. 2
Petrol Purple JR
Petrol Purple
Petrol Purple K1 *(Bluest)*

BROWNS

Petrol Brown V7 *(Yellowest)*
Petrol Mahogany L *(Reddest)*

BLUES

Petrol Azure Blue *(Reddest)*
Petrol Blue TX
Petrol Blue AA
Petrol Blue No. 2
Petrol Blue No. 3
Petrol Traffic Blue
Petrol Blue BB *(Greenest)*

GOLDS

Petrol Gold G *(True Gold)*
Petrol Gold P
Petrol Gold Leaf
Petrol Actec Gold *(Brownest)*

YELLOWS

Petrol Yellow C *(Greenest)*
Petrol Yellow G. T.
Petrol Yellow TT
Petrol Yellow WT *(Reddest)*

ORANGES

Petrol Orange Y *(Yellowest)*
Petrol Orange G
Petrol Orange WX
Petrol Brilliant Orange A
Petrol Orange P
Petrol Mandarine Orange P3 *(Reddest)*

REDS

Petrol Red P *(Orangest)*
Petrol Red D
Petrol Red C
Petrol Rubylite
Petrol Red NS *(Bluest)*

BLACKS

Petrol Black Q *(Bluest)*
Petrol Black A
Petrol Wax Black B *(Brownest)*

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Specify "Petrol" for Petroleum Colorings.

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An improved type of material for producing fluorescence in oils.

Bright Fluorescent shade, good yellow body with green fluorescence used in increasing fluorescent tone of lubricating oil.

100% soluble, faster to light than natural color in oil, stable to heat.

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FLUORESCENT ORANGE	A good bright Fluorescent shade useful for shading oils, produces a chalky-bloom.
DEBLOOMING AGENT A	Has the property of neutralizing blooms in lubricating oils.
N. P. A. COLOR	Has the property of changing body color of pale oils to red cast without affecting natural fluorescence.
COLOR SOLVENT #9	Increases brilliance and solubility of colors in waxes and oils.

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Oils	DETERGENTS
Rust	WETTING AGENTS
HIGH PRESSURE LUBRICANTS	DISPERSING AGENTS

The "Patent" laboratories and technical staff are available for the development of special colors or for the solution of any color problem.

Solvent Refining Processes

SOLVENT refining methods in the last few years have effected radical changes in the processing of lubricating oils, have been used extensively for the preparation of feed stocks for catalytic cracking, and have served in a variety of special applications for the separation of close-boiling hydrocarbons which could not be accomplished in conventional distillation apparatus.

In solvent refining, hydrocarbon mixtures are separated on the basis of their relative solubilities. Aromatic compounds, naphthenic compounds, and paraffinic compounds can be segregated without regard to volatility or the possibility of the thermal decomposition effects which often make distillations of heavy oils difficult.

In lubricant manufacturing the good properties of paraffinic materials are sought, and by the modern solvent refining processes oils of any desired quality can be uniformly produced from a wide variety of crude oils, with the percentage yield being the principal variable. The yield is directly dependent on the quantity of oil of the desired quality present in the original stock. No physical changes occur in the solvent refining processes, but only the separation of different groups of hydrocarbons. The development of the solvent refining processes has to a large degree eliminated the sharp distinctions once made between different crude oils on the basis of their suitability for lubricating oil manufacture. Now crude oils are evaluated more nearly on the amount of lubricating oil of specified quality which they will yield. The old practice of classifying crude oils as either suitable or unsuitable for high quality lubricating oil manufacture to a considerable extent has been abandoned.

Low carbon residue is desirable in feed stocks charged to catalytic cracking units, since the amount of carbon deposited on the catalyst in the cracking reaction is a principal factor in setting the throughput capacity of the unit. Several methods have been employed for preparing such low carbon residue feed stocks, including simple flashing, distillation, viscosity breaking, delayed coking, and solvent refining. Solvent refining is especially suitable for the purpose because in addition to the ability to prepare a very clean stock of good color properties, such methods also avoid temperatures that cause thermal decomposition of the feed stock. Flashing and distillation methods are likely to cause some decomposition since high temperatures are needed to vaporize heavy oils. The other feed preparation routes accept some thermal decomposition as an inherent part of the process method.

Besides these applications of solvent refining methods to lubricating oil manufacture and to deasphalting of cracking unit feed, there are many other cases where the separation of hydrocarbons by group types is essential and may constitute the only practical route for the manufacture of a product. One such case is the extraction of toluene from selected crude oils or from Hydroformer effluent. This separation could not well be accomplished by distillation because the components present have similar boiling points. During the war solvent extraction methods served to furnish most of our toluene requirements.

New developments in the manufacture of petrochemicals bring many more such cases into prominence. The plants currently under construction for the manufacture of synthetic gasoline are expected to have large volumes of oxygenated chemicals that will present complex fractionation and separation problems. It is expected that solvent separation methods will be found especially useful.

Duo-Sol Solvent Extraction Process

MAX B. MILLER & COMPANY, INC.

THE Duo-Sol Solvent Extraction Process is a method for the production of lubricating oils of desirable quality from any base stock, either residual or distillate.

The Duo-Sol Process is a system of extraction in which two partially miscible solvents of different specific gravity flow counter-current to each other, with the admission of the charge stock at an intermediate point in the system. The immiscibility of the solvents produces a system consisting of two layers of sufficient specific gravity difference to allow gravity separation in each of the exterior stages. The lighter layer is designated as the Raffinate layer and the heavier layer as the Extract layer. Propane, used as the raffinate solvent, is pumped into the Extractor No. 1 (E-1) together with the extract layer from Extractor No. 2 (E-2). Selecto, a blend of phenol and cresol, used as the extract solvent, is pumped into the Extractor No. 7 (E-7) together with the raffinate layer from Extractor No. 6. As can be seen from the flow diagram, the extract layers are pumped through the system and the raffinate layers flow by pressure differential, controled at the outlet of E-7 by a pressure control valve. While the solvents flow counter-current through the extraction system as a whole, the separation into layers in the individual extractors is obtained by flowing the mixture into a shallow box-type settling chamber extending the whole length of each extractor. The velocity, in the settling chamber, is maintained at less than turbulent flow rates to allow settling. The crude charge is pumped into Extractor No. 3 (E-3) together with the raffinate layer from E-2 and extract from E-4. This true counter-current extraction permits the propane to remove in stages E-1, E-2, all traces of raffinate material, resulting in the maximum yield of a given quality of oil.

The solvents Propane and Selecto are removed from the extract and raffinate layers by distillation in raffinate and extract evaporators P-1—P-2 and N-1—N-2, which consist of an upper section for high pressure propane evaporation, and a lower section of the same vessel for evaporation of Selecto at low pressure. Attention is drawn to the use of the latent heat of the selecto vapor to effect the evaporation of propane in towers P-1 and N-1 through the use of the external heat exchangers No. 19 and No. 20. The final traces of selecto are removed by a steam stripping in towers P-3, P-4, N-3 and N-4. This steam is removed from the distillation system as a constant boiling mixture at the top of tower W-1. After condensation and decantation the water containing Selecto is pumped to the propane charge tank to reduce the selecto content to approximately two percent; from here it flows to tower W-4 where advantage is taken of the fact that pure water boils at a slightly higher temperature than a Selecto-water constant boiling mixture. Steam overhead of W-4 containing Selecto is employed for vacuum jets, while the steam obtained from the steam drum (F-19), essentially Selecto-free, is employed after superheating for the final steam stripping in the vacuum towers P-4 and N-4, resulting in exceptionally low Selecto losses to the products.

The Duo-Sol process is the only solvent extraction process commercially employed for the production of a full line of motor oils by continuous operation on a single long residual charge stock, which after extraction, is continuously dewaxed andl rerun to final products of desired viscosity specification. Commercial Duo-Sol units have run continuously in this manner for as long as ten years on the same long residual charge stock, varying operation conditions to meet market requirements and interrupting operating only for periodic inspection and maintenance. The elimination of intermittent blocked operation and the drastic reduction in storage, testing and handling of many semi-finished stocks which the long residual processing permits, results in saving in both capital and operating costs.

The Duo-Sol process, however, is not restricted to the long residuum type of operation. Maximum flexibility is available to meet any requirement that changing market conditions might impose, due to the ability of the process to refine any distillate or residual stock. Distillate oils may be processed with very high yield and at a high through-put. In the case of Bright Stocks the Duo-Sol process is the only single process that can charge short residual stocks and produce premium oils of high viscosity index, low carbon residue, and extremely light and stable color. As an example of the flexibility of the process, in the space of a month and a half one commercial plant has treated satisfactorily a light and a heavy Colombian distillate having gravities of 21.2° A.P.I. and 19° A.P.I. respectively, an Oklahoma City distillate having a gravity of 28.8° A.P.I., and an Oklahoma City residuum having an A.P.I. gravity of 21.0°.

At the present time eleven plants are in operation in the United States with a total charge capacity of 32,600 bbls. per day. Two plants under construction will add 16,000 bbls. daily charging capacity to this total. In foreign countries eleven plants have been constructed with a capacity of 16,000 bbls. daily, ranging from 250 to 10,000 bbls. daily capacity.

Data from plant operation illustrating to a limited extent the flexibility and scope of Duo-Sol Solvent Extraction Process.

	East Texas-Louisiana Residuum	San Joaquin Long Residuum	East Texas Distillate	Van Zandt Long Residuum	Iraq Long Residuum
Charge Stock:					
Gr. (A.S.T.M.)	15.2	14.9	26.2	17.4	14.8
Vis. @ 210° F. (SUS)	760	257	47.4	179	194
V. G. C.	.876	.893	.839	.879	.899
Flash (°F.)	595	470	420	435	420
Raffinate—(Dewaxed):					
Gr. (A.S.T.M.)	26.5	29.1	29.9	27.5	27.6
Vis. @ 210° F. (SUS)	143.3	61.0	48.0	64.4	54.2
V. G. C.	.809	.808	.811	.818	.822
V. I.	95	94	95	98	100
Pour	0	0	0	0	0
Color (A.S.T.M.)	8	4¼	2¾	5	3½
Duo-Sol Yield percent by Vol.	43.6	34.7	73.0	60.9	55.0
Extract:					
Specific Gravity	1.026	1.0177	1.000	1.053	1.073

DUO-SOL SOLVENT EXTRACTION PROCESS
COURTESY MIX E MILLER & COMPANY INC
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
EXTRACT STRIPPER
EXTRACT EVAPORATOR
RAFFINATE EVAPORATOR
RAFFINATE STRIPPER
DEHYDRATOR
STRIPPING STEAM GENERATOR
HEATER N2,5,4
HEATER P2
STEAM
VENT
TO W-3
PROPANE
PROPANE CHARGE
CRUDE FEED
EXTRACT
RAFFINATE
EXTRACT LAYER
RAFFINATE LAYER
EXTRACTOR SYSTEM
350 - 400 PSI
100 - 150 °F
L.P. COMPRESSOR
H.P. COMPRESSOR
MIXER F4
SELECTO
SELECTO CHARGE

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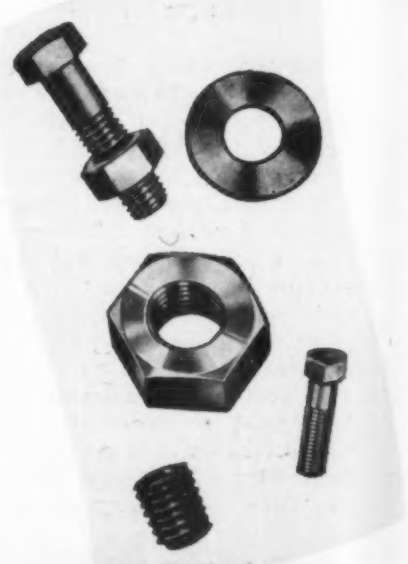

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Eliminate costly algae, slime, and odors for cooling water systems by installing a rugged and dependable Paddock Chlorinator. Available in 8 capacities ranging from 3-30 lbs. to 200-2000 lbs. of chlorine per 24 hours.

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Designed for the small water system. Operates by water pressure. Positive check on correct chlorine content is quickly and easily determined by a color density test. Built-in test set with each SURECLOR. Ideal as emergency stand-by unit for gaseous chlorinators.

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Nitrobenzene Extraction Process

THE ATLANTIC REFINING COMPANY

THE Nitrobenzene Extraction Process for separating the high grade lubricating oils, present in either the lube distillates or the residuum of any crude, from low grade components was developed and is operated commercially by The Atlantic Refining Company.

A crude oil lubricating distillate or residuum is a mixture of hydrocarbons, some highly paraffinic, some of intermediate character, and some highly naphthenic or asphaltic. The highly paraffinic compounds are practically insoluble in nitrobenzene, the asphaltic and highly naphthenic compounds are very soluble, while the compounds of intermediate character have solubilities between the two extremes. Hence, when oil is mixed with nitrobenzene the least desirable components are most readily dissolved and removed.

The process is physical in nature, the chemical character of the various hydrocarbons remaining unchanged. The maximum yield is equal to the amount of oil of desired quality originally present in the stock, and is closely approached in practice.

Nitrobenzene has a high specific gravity and a low viscosity at the temperatures employed for extraction. As a result, mixing and settling in the extractor system are rapid and complete. The boiling point is low enough to permit a complete separation of solvent and oil by simple evaporation and steam stripping. Nitrobenzene and water are practically immiscible and the separation of the 0.2 percent of solvent dissolved in water is very simple. At operating and storage temperatures the solvent has low volatility, so that losses are small.

Nitrobenzene is distinctive in its high solvent power for asphaltic, naphthenic, and aromatic compounds. The resultant low solvent-oil ratio to produce a given yield of a given quality oil, varying from 0.5 volumes to 2 volumes of solvent per volume of oil, is important since solvent recovery is the major cost item in solvent refining.

Essentially the process consists of two steps: contacting the charge oil with nitrobenzene in a multitage countercurrent extractor, composed of a series of mixers and settlers; and then recovering the nitrobenzene from the raffinate and extract solutions. The raffinate solution contains the paraffinic fraction of the charge stock in which some nitrobenzene has been dissolved, while the extract solution consists of nitrobenzene and the undesirable material which it has dissolved. The nitrobenzene recovered from these solutions is re-used.

The flow of material is illustrated by the diagram on the opposite page. The charge oil is first pumped through scraper water-coolers and thence to scraper brine-chillers. At this point it is diluted with extract solution from the second stage of the extractor. The mixture is then cooled to the operating temperatures and enters the first settler. Nitrobenzene flows through an exchanger and cooler to enter the extractor at the fifth stage mixer, and flows through the extractor system counter-current to the oil. The raffinate and extract solutions flow continuously from each settler into mixers or to the recovery system. The extract solution flows from extractor 5 to extractor 1 in succession, while the raffinate solution flows from 1 to 5. The extract solution leaving extractor 1 and the raffinate solution leaving extractor 5, in which it has been contacted with pure nitrobenzene, are ready for solvent recovery. The extractor system may be operated at constant temperature (40-80° F.) throughout or with a temperature gradient, as conditions dictate.

The extract solution is heated in a series of exchangers and then delivered to the evaporator-condensers B, where the nitrobenzene present is boiled off under reduced pressure, using closed steam as the heating medium. The nitrobenzene vapor is condensed for re-use. Plant operating pressures do not exceed atmospheric, and the maximum temperature used is 360° F. The extract leaving the evaporators contains about 5 percent nitrobenzene which is removed by stripping with open steam in tower C. The raffinate solution is treated in a similar fashion in the evaporator-condensers D and stripper E.

Any water which has been in contact with nitrobenzene in the process is collected in settler F. The nitrobenzene which separates is dehydrated in evaporator H, and returned to storage. The aqueous layer from settler F is heated by exhaust steam to about 125° F. and about 15 percent of the water is vaporized under vacuum in evaporator G. The vapor, which contains all of the nitrobenzene in the water charged to evaporator G, is condensed and returned to the separator.

In general, oil of any quality up to Pennsylvania grade or better can be obtained from any distillate or residuum which contains any oil of this quality. The quality of the lubricating oil obtained is dependent on the degree of treatment.

Charge stocks varying from light distillates (150 S.S.U. viscosity at 100° F.) to heavy residuums (950 S.S.U. viscosity at 210° F.) have been successfully processed at an average rate of more than 3000 barrels per calendar day in the plant at Philadelphia. The color reduction obtained ranges from 60 to over 90 percent. The yield depends on the composition of the charge stock and the quality of product desired. The quantities of nitrobenzene required vary from one-half volume per volume of charge oil, rarely exceeding two volumes per volume of charge oil.

A typical waxy Mid-Continent S.A.E. 30 lube distillate would be processed at a 3500 barrel per stream day charge rate with a 1:1 solvent-to-oil ratio at 50° F. extractor temperature and the operating conditions given in the flow diagram. Such a charge stock 25.5° A.P.I. gravity, 4¾ Dilute A.S.T.M. Color, 75 S.S.U. viscosity at 210° F., and 0.828 V.G.C.) would process to a 75 percent raffinate yield of 30.2° A.P.I. gravity, 3½ A.S.T.M. color, 63 S.S.U. viscosity at 210° F., 0.798 V.G.C., and 100 V.I. after dewaxing to 0° F. Pour. The extract oil would test 13.0° A.P.I., 146 S.S.U. viscosity at 210° F., and 0.919 V.G.C.

The paraffinic oil made will have a higher A.P.I. gravity, a better V.G.C. and V.I., a much lighter color and lower carbon residue than the charge stock. The extent to which these improvements are carried depends on the degree of treatment used.

NITROBENZENE EXTRACTION PROCESS
Conditions shown are for a typical waxy Mid-Continent SAE 30 Lube Distillate
COURTESY ATLANTIC REFINING COMPANY
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
PRESSURE, MM HG ABSOLUTE
CRUDE CHARGE
WATER SCRAPER COOLER
SCRAPER BRINE CHILLERS
BRINE
COUNTER-CURRENT EXTRACTOR
EXTRACT STRIPPER-C
EXTRACT EVAPORATOR CONDENSERS
NITROBENZENE DEHYDRATOR-H
VACUUM JET
NITROBENZENE RECOVERY FROM WATER-G
RAFFINATE STRIPPER-E
NITROBENZENE SEPARATOR-F
STEAM
EXTRACT SOLUTION
RAFFINATE SOLUTION
NITROBENZENE SURGE
RAFFINATE OIL TO DEWAXING
EXTRACT OIL

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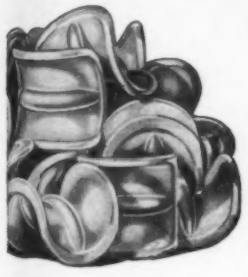
Tower Packing

Knight-Ware Tower Packings are widely used in scrubbing, drying, recovering, reaction, rectifying, extraction and distillation columns. Popular shapes include Berl Saddles, Raschig Rings, Partition Rings and Spiral Rings. They are usually supplied unglazed since the natural matte surface wets better than a glassy surface.



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Berl Saddles

Our most popular tower packing is Berl Saddles. They provide from 20 to 50% more usable surface and up to 100% less resistance per unit volume. These properties make them especially suitable for vacuum distillation, extraction and high velocity columns. Besides permitting smaller sized equipment, in many instances they will pay for themselves with savings on power costs alone. Other advantages are: minimum channeling, reduced side thrust, and a stronger angle of distribution. Made of Knight-Ware or porcelain, Berl Saddles are corrosion-proof.

All Knight-Ware Tower Packing is made from selected, washed and de-aired clays. Each type will withstand severe acid service, has a high crushing strength and will not spall. The one-inch Raschig Rings, for example, have shown an average of 125 lbs. in break tests.

BERL SADDLES

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¼	113,000	60	274	58
½	17,600	56	155	60
¾	6,000	49	99	65
1	2,300	44	79	69
1½	690	42	52	70

RASCHIG RINGS

¼	88,000	67	220	52
⅜	26,000	65	148	53
½	10,700	65	114	53
⅝	5,800	64	80	54
¾	3,000	46	72	67
1	1,330	45	58	68
1¼	670	38	44	73
1½	380	45	36	68
2	162	37	28	75
3	48	40	19	74

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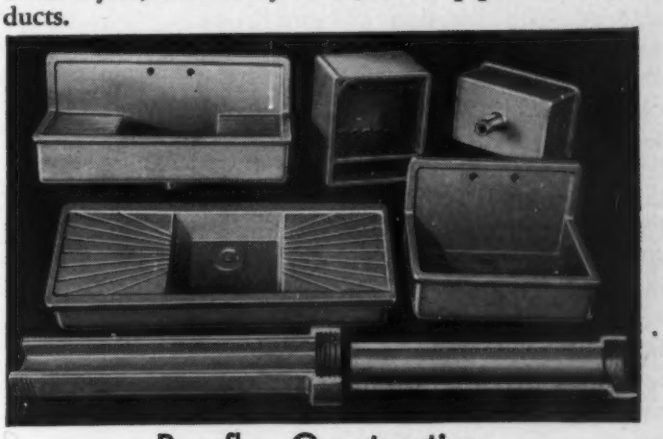
Some of the forms in which Knight-Ware equipment is available, in standard or special designs, are: Pipe and fittings in bores of from 1″ to 60″; towers in diameters of from 8″ to 60″; acid jars, filters and kettles, 5 to 200 gals.; plug cocks in plain and lubricated styles; laboratory sinks, drain pipe and fume ducts.



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Pyroflex is an acid- and alkali-resisting lining applicable to steel, concrete, wood or masonry shells. This non-porous, flexible, seamless lining will not deteriorate with heat or age. It may be applied on the job or at our factory.

Knight-Ware brick or tile over the Pyroflex provides an extremely rugged acid- and leak-proof combination lining. Carbon brick may be substituted for Knight-Ware to handle hydrofluoric acid and caustic soda.

We have installed linings or furnished completed units for such service as: steel pickling, metal plating, electrolytic refining, acid storage, acid concentration, salt crystallizers, acid trenches and neutralizing pits; equipment such as gas scrubbers, absorption towers, chlorine bleach towers and acid fume eliminators.

Please furnish sketches and state service conditions when requesting quotation.

Pyroflex units such as fume washers, cooling towers, absorbers and concentrators, are individual, custom-built units designed to meet the exact needs of the customer.

Furfural Refining Process

TEXACO DEVELOPMENT CORPORATION

FURFURAL is an extractive solvent extensively employed for refining a wide range of lubricating oils of any viscosity and diesel-fuel oil. Furfural originally was chosen as most suitable for solvent extraction of lubricating oils because of its high degree of selectivity for the undesirable components of petroleum oils without possessing a high solvent power for the desirable components. It is unique in this respect. The addition of water or an "opposing" solvent is not required to control this solvent power. Furfural also was chosen because of its stability, low cost, availability and desirable physical characteristics. Being non-acidic and non-toxic, furfural can be handled without special protective provisions. Its boiling point is sufficiently low (323° F.) to permit its recovery from the lowest-boiling-point lubricating-oil stocks and even from diesel-fuel oil or furnace oils without appreciable loss of oil or solvent. Its low solid point (—38° F.) eliminates the danger or solidification in lines or tanks in cold weather. The high specific gravity of furfural insures rapid and complete separation at normal operating temperatures in extraction towers of relatively small diameter. The critical solution temperature of furfural is high enough to permit its application to waxy or viscous oils at temperatures that insure treatment in a non-viscous, completely liquid state. The stability and recoverability of furfural is emphasized by the low operating losses experienced by commercial refining units.

The wide application of furfural to solvent refining is illustrated by the fact that units ranging from 60 to 6000 barrels per day of charge oil capacity, totaling over 41,000 barrels per day, have been installed in some 20 refineries for the solvent extraction of naphthenic or paraffinic distillate and residual oils. Additional furfural-extraction units having a combined charge capacity of over 40,000 barrels per day are under contract for construction.

The furfural-extraction unit represented by the accompanying flow diagram comprises an extraction or treating section, a raffinate-recovery section, an extract-recovery section, water-removal facilities, process surge tanks and pumping equipment. Furfural, however, because of its stability over a wide range of operating conditions, lends itself to a variety of methods of recovery from raffinate and extract solutions. Different types of the solvent-recovery systems have been developed and used, in order to provide for maximum operating economy of each particular installation.

One of the major improvements in the process in recent years has been the development of a highly efficient treating tower, in which an increased degree of fractionation is obtained between the relatively insoluble, desirable lubricating-oil fractions and the more soluble, undesirable fractions. Untreated oil is introduced into the continuous countercurrent extractor at some mid-point and at a predetermined temperature of from 110° to 200° F. The temperature of the furfural charged near the top of the extractor varies depending on the charge stock, up to 290° F. or higher. A substantial temperature gradient is maintained between the top and bottom of the tower by indirect water coolers provided to control the temperature spread.

The extraction tower generally is operated with a prevailing oil phase through which the dispersed furfural phases passes downward over Raschig ring nests which afford an extensive surface of contact between the phases. This, in turn, promotes rapid and relatively complete equilibrium by interchange between the phases without mechanical mixing or pumping equipment. The concentration of undesirable oily constituents in the furfural phase increases toward the bottom of the tower from which the extract solution is withdrawn continuously. Conversely, the oil quality improves toward the top of the tower from which practically all oil of predetermined raffinate quality that is present in the charge stock, overflows with a small percentage of furfural dissolved in this phase.

Further improvement in fractionation between desirable and undesirable constituents has been secured by the controlled recirculation or "refluxing" of extract to the lower portion of the tower. The recirculated extract functions in much the same way as reflux to a distillation column, in that it sharply fractionates oil from the extract solution, and this oil rises again to the refining zone for re-extraction and consequent improvement in raffinate yield.

The recovery of solvent is accomplished by continuous distillation of furfural from the raffinate and extract phases in the respective recovery sections followed by steam stripping of each product. The overhead condensate from the steam strippers separates into two furfural-water layers, one rich and the other lean in furfural. The recovery of furfural from water is accomplished in a simple continuous fractionating system in which dry furfural is removed from the bottom of one tower and water from the bottom of another tower, taking overhead from each a constant-boiling mixture of water and furfural which is condensed and returned for redistillation.

This discussion is presented through the cooperation of Texaco Development Corporation, 135 East 42nd Street, New York, New York, licensors of the Furfural Refining Process.

FURFURAL REFINING UNIT
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FURFURAL-FROM-WATER STEAM STRIPPER
EXTRACT MIX HEATER
REFINED OIL MIX HEATER
HOT FURFURAL ACCUMULATOR
LEGEND
VAPOR HEAT EXCHANGER
SOLVENT-EXTRACT MIX EXCHANGER
WATER-FROM-FURFURAL STRIPPER
FURFURAL STORAGE
EXTRACT ATMOSPHERIC FLASH TOWER
EXTRACT PRESSURE FLASH TOWER
REFINED OIL STEAM STRIPPER
EXTRACT STEAM STRIPPER
EXTRACT COOLER
EXTRACT STORAGE
LOW PRESSURE REFINED OIL FLASH TOWER
REFINED OIL-REFINED OIL MIX EXCHANGER
REFINED OIL COOLER
FURFURAL REFINED OIL STORAGE
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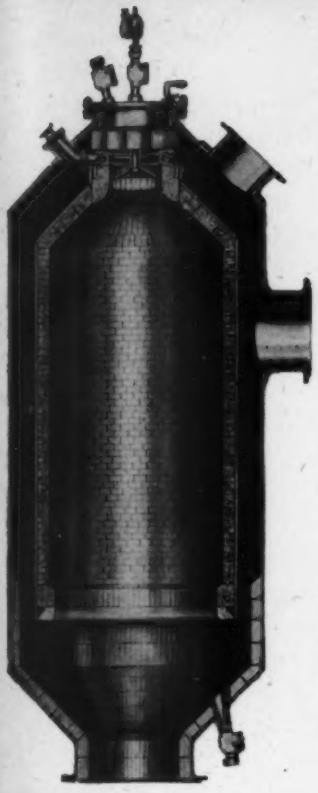

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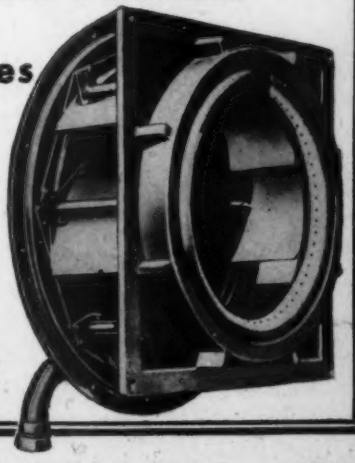

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PEABODY LIMITED, LONDON

Phenol Extraction Process

THE M. W. KELLOGG COMPANY

PHENOL extraction is a liquid-to-liquid extraction process for improving the viscosity index and oxidation resistance, and for reducing the carbon and sludge-forming tendencies of lubricating oils made from crude petroleum.

The process is employed successfully either before or after dewaxing on both distillate and residual oils from all types of crudes. The selective solvent employed—phenol—gives high yields of desired products and has exceptional stability. The solvent is continuously re-used with losses approximating 0.01 percent of the total solvent circulated per day in well-operated plants. The high solvent power of phenol necessitates only comparatively low ratios of solvent to oil, a feature which is important in securing low operating costs.

A total of 11 phenol extraction plants now operating in the United States and Canada have a combined oil-charge capacity of about 45,000 barrels per day. Six additional units now being installed in the United States will have a combined oil charge of about 23,000 barrels per day.

Modern phenol extraction plants are continuous throughout and employ a very efficient vertical countercurrent extraction tower, the design of which permits usually high oil-charge rates per square foot of cross section. Heat requirements are kept to a minimum by utilizing heat from the condensing phenol vapors to vaporize solvent going to the extract-recovery system.

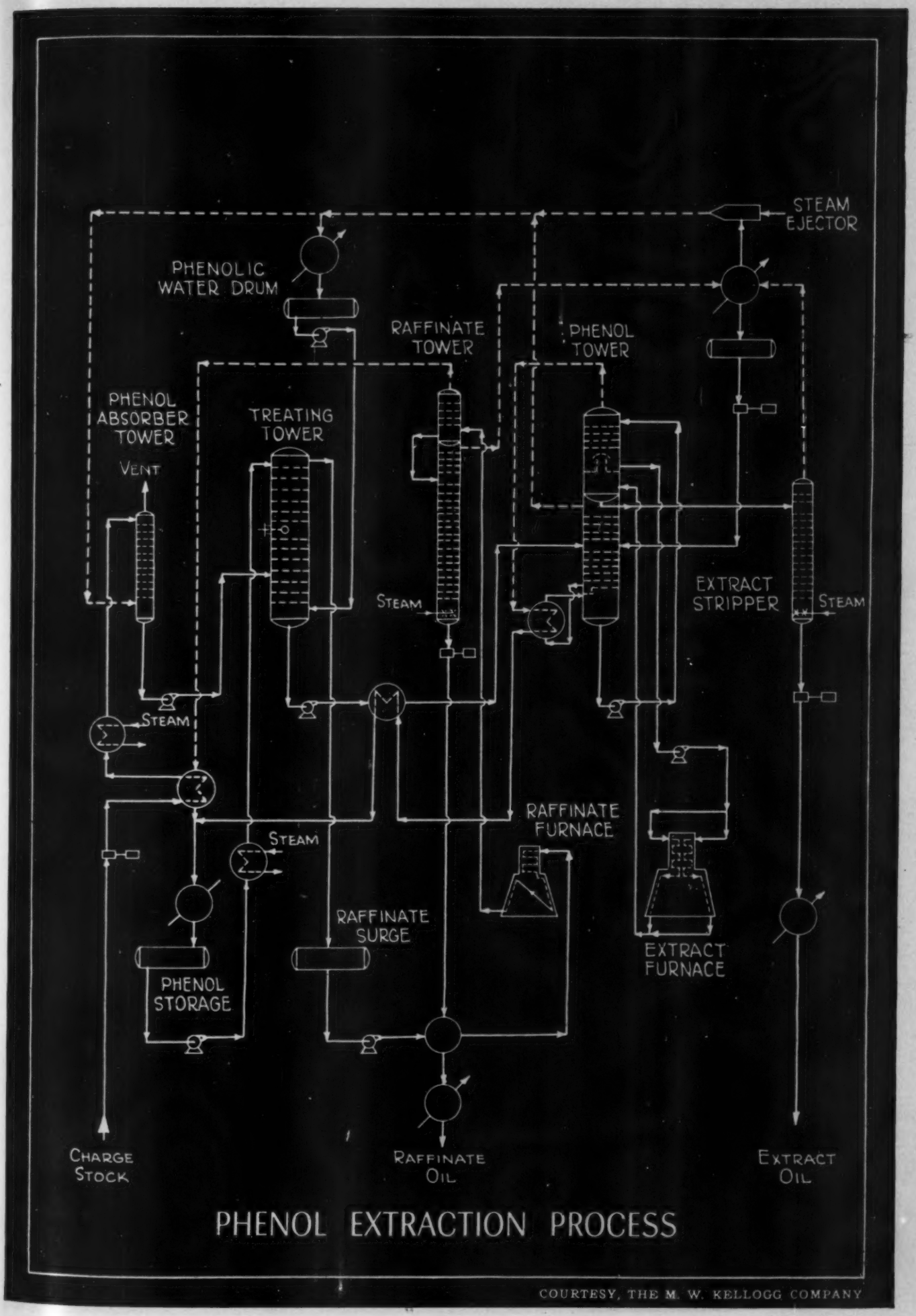
A simplified flow diagram, outlining the complete treating and solvent recovery systems, is shown on the opposite page. Raw stock and dry phenol are fed into the treating tower. Prior to entering the treating tower, the feed stock flows through the phenol absorber tower to remove the remaining phenol from process water being discharged from the plant.

Raffinate phase is withdrawn from the top of the treating tower to the raffinate storage drum from which it is fed through a heat exchanger, exchanging heat with the treated raffinate oil, thence through a coil in the raffinate furnace and is flashed into the evaporator section of the raffinate tower at substantially atmospheric pressure. Phenol vapor recovered in this section flows to the phenol condenser and enters the phenol storage tank.

The raffinate oil flows to the stripper section of the raffinate tower where it is stripped free of solvent with steam and is withdrawn from the bottom to storage. Steam and solvent vapors flow from the top of the stripping section to a surface condenser and steam jet, the purpose of which is to maintain a suitable vacuum for the stripping operation. The extract phase is withdrawn from the bottom of the treating tower and is fed through heat exchangers, in which heat is recovered from phenol vapors, and enters the drying section of the phenol tower in which all water is removed as constant-boiling phenolic water vapor. The phenolic water vapor leaving the top of this section is condensed and enters the phenolic-water drum from which is drawn reflux for the drying section, as well as the water required for modification of solvent at the bottom of the treating tower. Dry extract solution is pumped from the bottom of the drying section of the phenol tower to the top of the evaporator section of the phenol tower where a portion of the solvent is flashed off. Phenol vapor reovered in this section flows through exchangers into the phenol-storage drum. The remaining extract solution drops to a pan in the evaporator section of the phenol tower and is pumped through a coil in the extract furnace returning to a point below the pan in the evaporator section of the phenol tower. Extract oil containing a small amount of dissolved phenol flows from the bottom of the evaporator section of the phenol tower to the extract stripper where the last traces of solvent are removed by a stream-stripping operation, similar to that used for recovery of phenol from raffinate. Stripper extract is withdrawn from the bottom of this tower to storage.

Treating results obtainable with phenol extraction may be gauged from the table below summarizing yields, treating conditions and product qualities for several different stocks.

	Deasphalted East Texas Short Residue	East Texas Neutral	West Edmonds SAE 40 Stock	Deasphalted Oklahoma City Residue
Solvent Quantity, Vol. Percent	300	128	213	227
Treating Temperature, °F	245	168	201	200
Raffinate Yield, Vol. Percent	64.5	74.1	88.2	82.2
Inspections Untreated Oil:				
°API Gravity	23.0	26.2	28.3	23.8
SSU Vis. at 210 °F	159.0	44.5	77.4	156.6
Color—NPA	7—	2½ D	3—	4½
Carbon Residue, Percent	1.96	....	0.16	0.73
Flash COC, °F	590	400	530	595
Fire COC, °F	680	455	595	665
V. G. C.	.833	.842	.808	.828
Dewaxed (Zero Pour) V. I.	59	63	81	74
Raffinate Oil:				
°API Gravity	29.0	32.6	30.2	27.1
SSU Vis. at 210 °F	110.7	42.3	72.3	137.6
Color—NPA	5—	1½—	2—	3½—
Carbon Residue, Percent	0.26	....	0.04	0.24
Flash COC °F	600	405	530	605
Fire COC °F	660	455	595	675
V. G. C.	.795	.800	.796	.805
Dewaxed (Zero Pour) V. I.	94	93	93	93



PHENOL EXTRACTION PROCESS

COURTESY, THE M. W. KELLOGG COMPANY

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Edeleanu Process

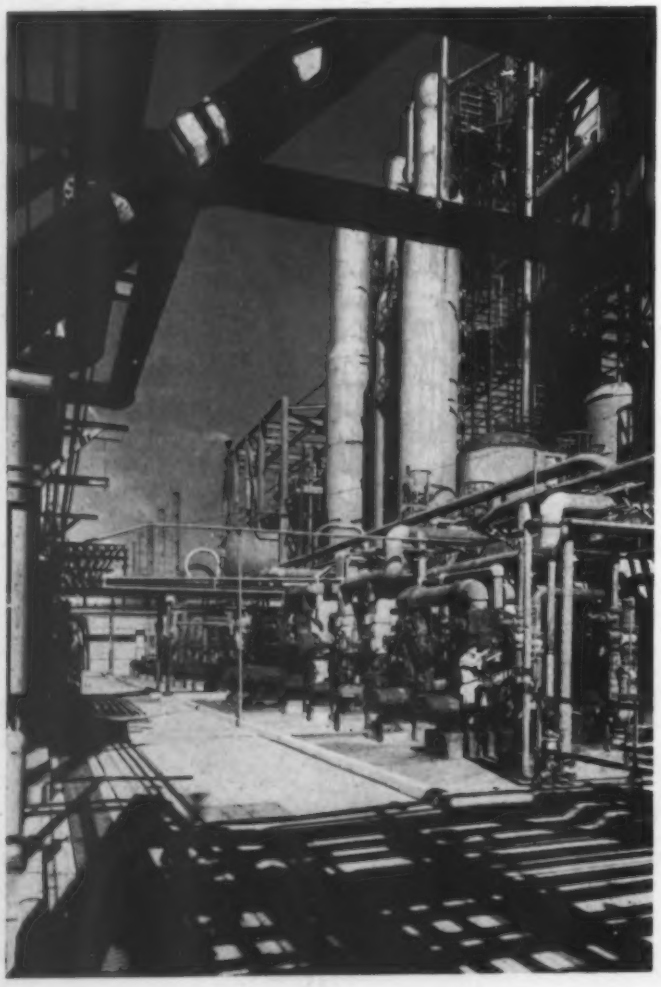
E. B. BADGER & SONS COMPANY

THE Edeleanu Process, employing liquid sulfur dioxide as a solvent, has been used commercially since 1911 for refining petroleum products. Its first application was for the production of high-grade kerosines from highly aromatic and sulfur-bearing crude kerosine distillates. This advance over heavy acid treatments with their attendant high loss of hydrocarbons led to its eventual application to the entire range of petroleum products. The Edeleanu Process is unique insofar as it is the only solvent refining process by which all fractions of mineral oils may be extracted. The low boiling point of sulfur dioxide (14° F.) makes the problem of recovering the solvent relatively simple even though hydrocarbons in the gasoline boiling range are treated.

The flow diagram on the opposite page, presented through the cooperation of E. B. Badger & Sons Company, 75 Pitts Street, Boston, Massachusetts, shows the essential equipment in a typical Edeleanu Unit. A study of the flow sheet shows that the plant consists of the following main sections:

(1) Counter-current extraction tower for contacting oil and solvent.

(2) Equipment for the recovery of the small amount of solvent which dissolves in the refined oil.



(3) Equipment for recovery of the bulk of the solvent which leaves the extraction equipment in the extract.

(4) Equipment for cooling oil and solvent, compressing, recondensing and dehydrating the solvent.

Conventional equipment is employed and a simple system for continuously dehydrating the liquid sulfur dioxide eliminates corrosion. With minor exceptions, ordinary steel equipment is used throughout the plants.

Liquid sulfur dioxide preferentially dissolves aromatic hydrocarbons. It has a less definite preference for naphthenic hydrocarbons and little solubility for paraffines. The degree to which naphthenes and paraffines are dissolved depends upon the temperature of extraction and the boiling range of the stock being treated. The lower the extraction temperature, the less paraffines and naphthenes will be dissolved. The lower the boiling range of the material being extracted, the greater will be the solubility of the SO_2 for paraffines and naphthenes. Generally, therefore, lower-boiling materials are treated at lower temperatures than higher-boiling ones.

The extraction by liquid sulfur dioxide of naphtha is commercially accomplished at temperatures varying from +10° F. to —60° F. The raffinates will be practically free of aromatics and frequently are sold as special solvents (VM&P naphthas, Stoddard solvent, etc.) or are reformed. The extract generally will have an aromatic content of from 55 to 85 percent, depending on the temperature employed and will be of high octane number and of high solvency power. By the addition of another processing step the aromatic content may be increased to over 99 percent. This form of the process has been used to produce pure toluene from petroleum fractions.

A large portion of the installed capacity in Edeleanu units is used for the production of high-quality kerosines from aromatic and high-sulfur crudes. These operations generally are conducted at 10 or 20° F. In this use of the process not only is a kerosine raffinate of excellent color and burning characteristics produced, but the aromatic extracts also frequently are sold at a premium.

The widely increased use of diesel engines has stirred interest in the Edeleanu Process for the production of better diesel fuels. Higher cetane numbers and lower carbon residues and sulfur contents are obtained in SO_2 raffinates produced either from straight-run or from cracked gasoils.

A further indicated use of the process is for the extraction of catalytic cycle stocks. The SO_2 raffinate is substantially as good a cracking stock as the original virgin-gasoil charge. The extract produced is a highly aromatic and refractory material which will crack only to a very low yield of gasoline and is best diverted directly to fuel. The use of the SO_2 extraction step will result in obtaining highest overall gasoline yield with the least use of cracking-plant capacity.

EDELEANU TREATING PROCESS
COURTESY, E. B. BADGER & SONS COMPANY
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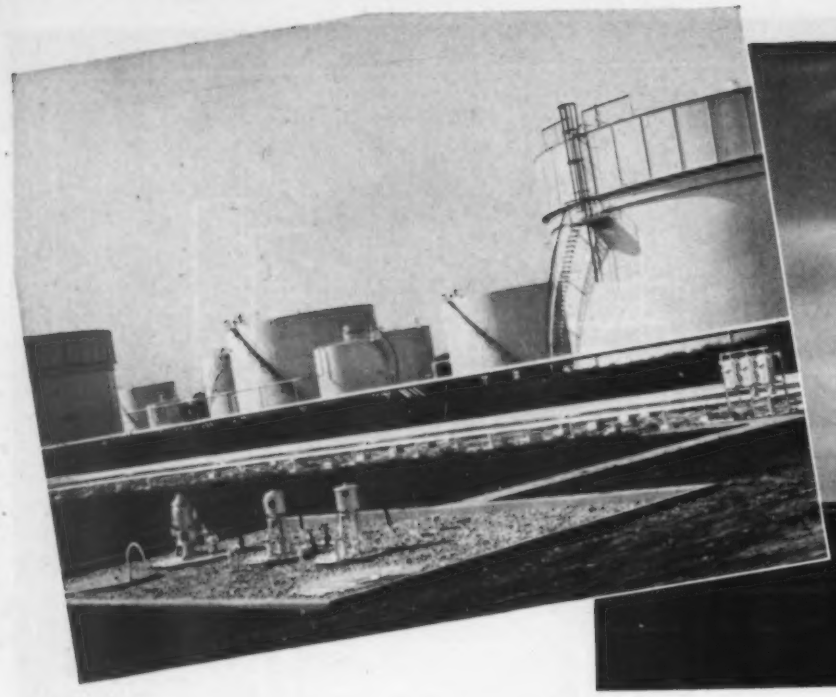
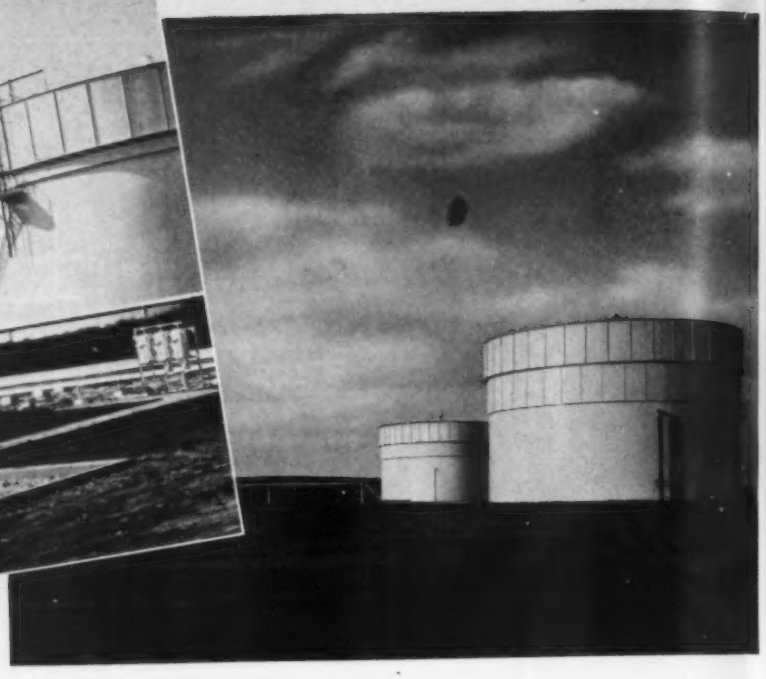
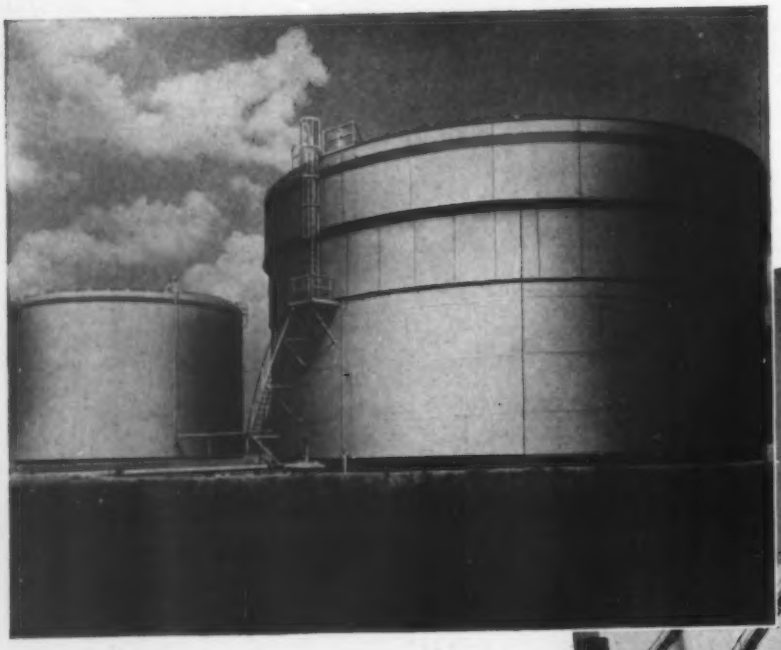
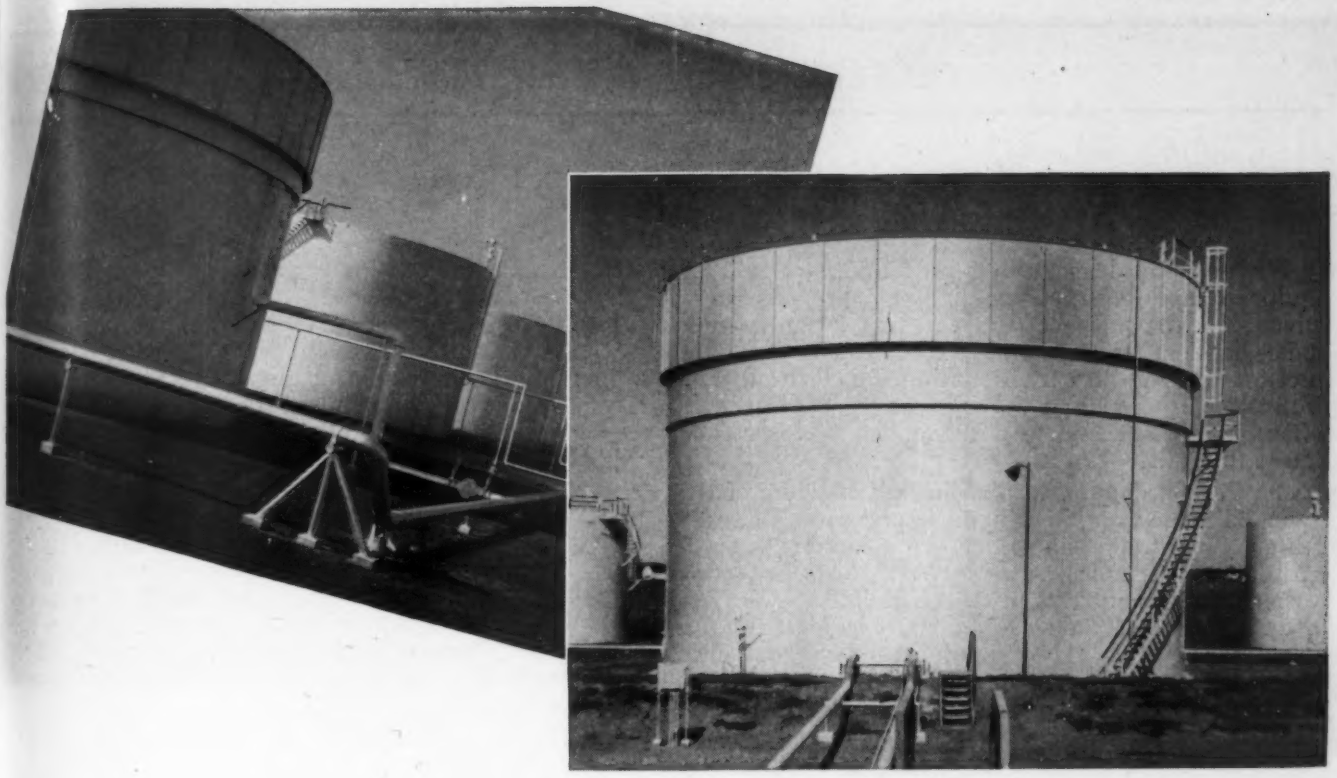
CHARGE FILTERS
CHARGE PRECOOLER
CHARGE CHILLER
EXTRACTION TOWER
SO2 CONDENSERS
MAIN SO2 COLLECTOR
SO2 COMPRESSOR
SO2 STORAGE
HIGH PRESSURE SO2 COLLECTOR
AIR-SO2 - EXTRACT SOLUTION EXCHANGER
EXTRACT A.P. EVAPORATOR
EXTRACT CONDENSER-RE PRESSURE EVAPORATOR
EXTRACT VACUUM EVAPORATOR
EXTRACT COMBINED PRESSURE EVAPORATOR
EXTRACT AIR STRIPPER
SOLVENT-WATER FRACTIONATOR
RAFFINATE CONDENSER-RECOVERY EVAPORATOR
RAFFINATE C.P. EVAPORATOR
RAFFINATE VACUUM EVAPORATOR
SO2 VACUUM PUMP
REFLUX
STM.
WASTE
RAFFINATE OUT
EXTRACT OUT





More EXPANSION





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Propane Deasphalting Process

THE M. W. KELLOGG COMPANY

THE propane deasphalting process is a liquid-to-liquid extraction process employing propane as a solvent for the purpose of separating asphalt, color bodies, and other undesirable constituents from reduced crudes, and for fractionating the clean oil thus produced into portions having different viscosities which may be subsequently further refined to finished lubricating oils.

The early propane deasphalting plants were of the so-called stage type, in which the reduced crude was contacted with propane in only one or two stages without appreciable counter-current action. These earlier stage-type plants made possible the separation of a considerable amount of hard asphalt from reduced crude but did not permit the production of deasphalted oils having light colors and low carbon residues suitable for subsequent refining with single selective solvents. Deasphalting plants installed in recent years have all been of an improved type with the older stage-settlers replaced by a vertical baffled tower. By employing a countercurrent flow of oil and propane within the tower and by properly controlling the temperature of the solution in the different portions of the tower, reduced crude is accurately fractionated in a series of stages into clean, light-colored deasphalted oil and high-melting point, relatively oil-free asphalt. It is probable that all future propane deasphalting plants will be of the tower type and it now appears that this process will come into general use as the primary step in manufacturing high-viscosity lubricating oils from reduced crudes.

The ability of the process to produce oil of excellent color and low carbon residue is completely independent of the asphalt content of the reduced crude. This fact, coupled with a low operating cost, makes it possible to produce lubricating oils commercially by this process from reduced crudes which have previously been considered unsuited for the purpose. Reduced crudes containing as much as 50 percent asphalt are presently being handled by this process and light-colored, low-carbon residue deasphalted oils are being produced therefrom.

Although the propane deasphalting process has been employed to date primarily for the purpose of separating reduced crude into asphalt and a deasphalted oil fraction, the process is also employed for the fractionation of the deasphalted oil into two or more portions. By a simple variation of operating conditions, the highly viscous oil portions of reduced crudes may be separated into as many different fractions as is desired and each fraction subsequently utilized for the purpose for which it is best suited. Because all separations thus made with propane are carried out at relatively low temperatures, no cracking occurs and the fractions secured are free from thermal decomposition products.

A simplified flow diagram of the propane deasphalting process is shown on the opposite page. Reduced crude is pumped from storage through a steam heater into the upper portion of the deasphalting tower. Propane is pumped from the propane-storage tank into the lower portion of the deasphalting tower from where it flows upward countercurrently to the descending asphalt solution phase. Deasphalted oil solution is withdrawn overhead and flows to a low-pressure steam propane evaporator, where a large amount of propane is flashed overhead through a condenser into the propane-storage tank. The deasphalted oil-propane solution flows from the low-pressure-steam propane evaporator to the high-pressure-steam propane evaporator where the majority of the remaining propane is flashed off, joining the stream from the low-pressure-steam propane evaporator. The oil flows from the high-pressure-steam evaporator into a steam stripper, where the remaining propane is removed. Deasphalted oil is withdrawn from the bottom of the stripper and pumped to storage.

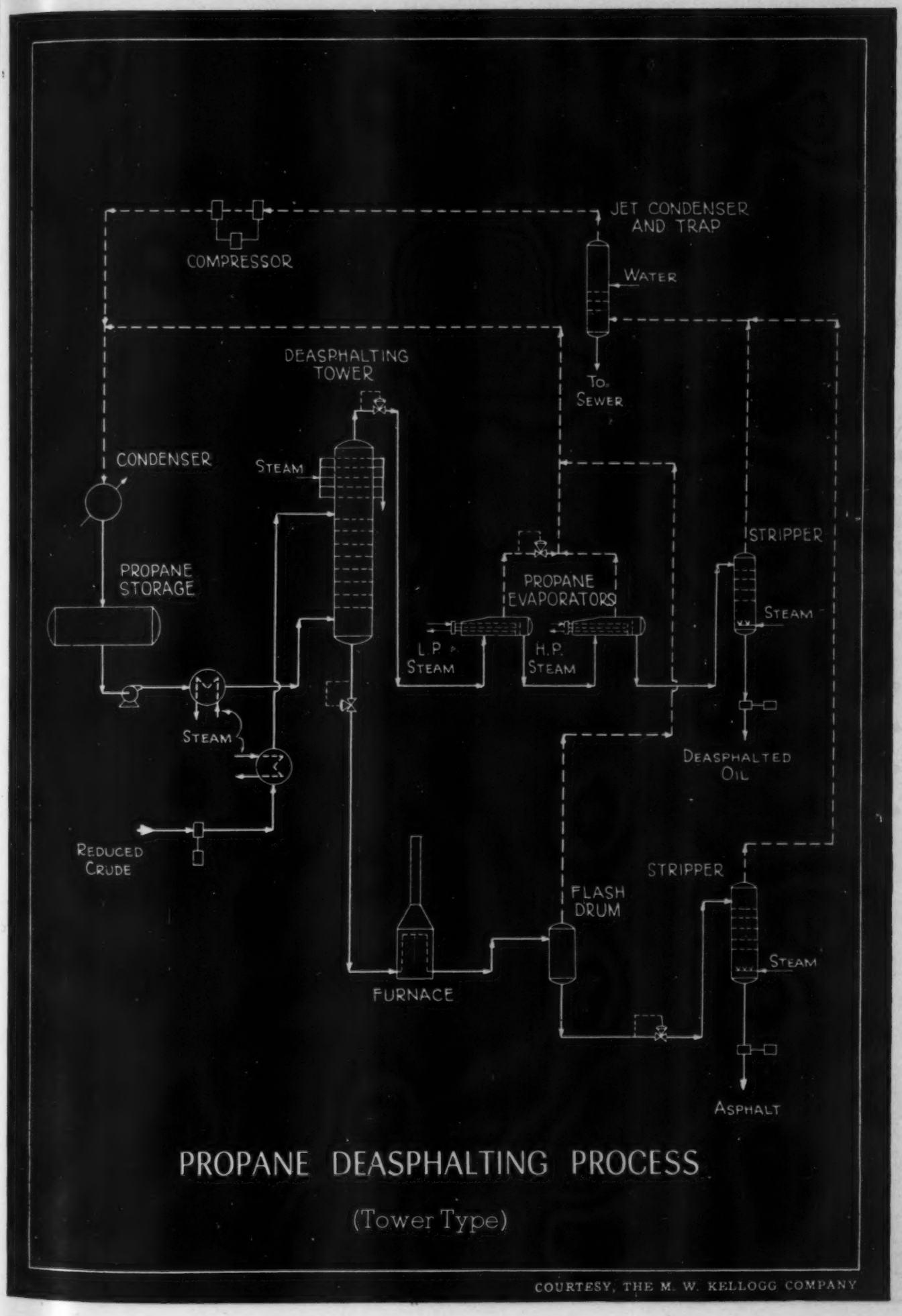
The asphalt solution is withdrawn from the bottom of the deasphalting tower and passes through a furnace and thence to a flash drum where the greater part of the propane is flashed overhead, joining the combined propane streams from the low-pressure- and high-pressure-steam propane evaporators. The asphalt with remaining small percentage of propane is withdrawn from the bottom of the flash drum into a steam stripper, where the remaining propane is removed. The asphalt product is pumped from the bottom of the stripper to storage.

The stripping steam and propane vapor from the asphalt and oil strippers passes to the jet condenser where the steam is condensed. The propane vapor goes to the compressor and thence to the propane condenser and storage tank.

Typical Yields—Tower Propane Deasphalting

	Van Zandt	California	East Texas	Heavy Mid-C.
Reduced Crude:				
°API Gravity	12.9	8.6	15.2	12.2
SSU Vis. at 210 °F	1249	7685	513	1716
Car. Res., Percent	13.9	12.1	10.9	15.2
Flash, COC °F	585	600	575	630
Fire, COC °F	675	690	665	695
V.G.C.	.890	.....	.882	.892
Deasphalted Oil:				
Vol. Percent Yield	58.8	32.8	63.9	54.0
°API Gravity	21.7	21.4	23.6	23.3
SSU Vis. at 210 °F	171.9	176.4	112.9	150.8
Color, NPA	6	8—	6	5—
Car. Res., Percent	1.81	1.73	1.32	1.55
Flash, COC °F	580	545	.....	595
Fire, COC °F	655	650	.....	655
V.G.C.	.842	.845	.836	.840
Asphalt:				
Vol. Percent Yield	41.2	67.2	36.1	46.0
Specific Gravity	1.0583	1.0445	1.0583	1.0603
Melting Point, °F	164	162	165	178

The M. W. Kellogg Company has been appointed licensing agent for the process for the United States and foreign countries. Licenses granted are under the patents and patent rights of Standard Oil Company (New Jersey), Union Oil Company of California, Standard Oil Company (Indiana) and The M. W. Kellogg Company.



PROPANE DEASPHALTING PROCESS

(Tower Type)

COURTESY, THE M. W. KELLOGG COMPANY

BORN "UPFLO" HEATERS

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With rising fuel costs it is more important than ever before to equip your plant with heaters operating at the highest efficiencies . . . . that calls for BORN UPFLO HEATERS.

Letters and reports from our customers consistently show over-all efficiencies ranging from 75% to 87% depending on type of installation and operating conditions.

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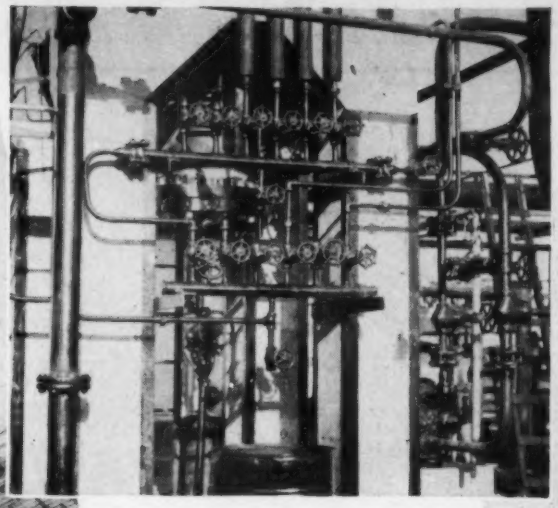

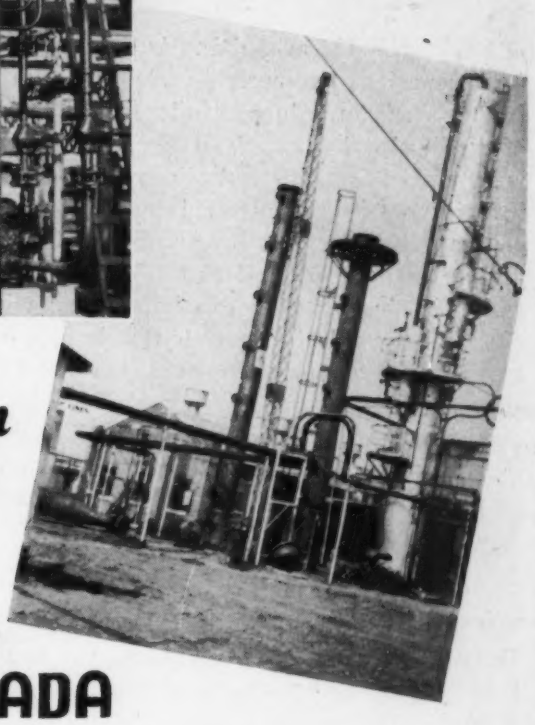

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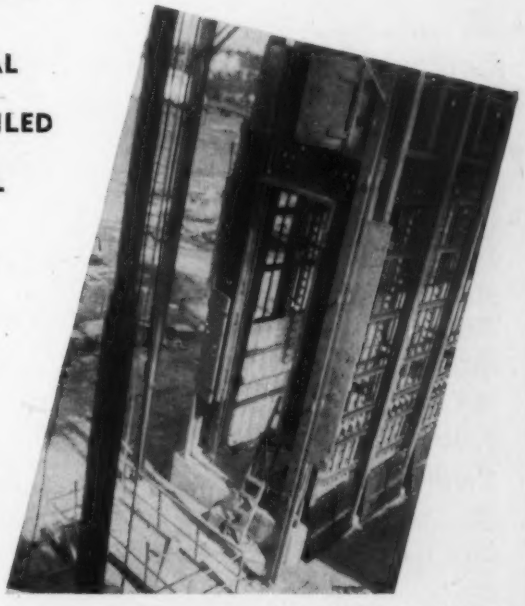

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Propane Decarbonizing Process

THE M. W. KELLOGG COMPANY

PROPANE Decarbonizing is a continuous process for preparing low-carbon-forming ctalytic-cracking feed stock from reduced crude. It is the exact reverse of distillation in that the highest-boiling component is separated first. Instead of relying upon a difference in boiling points, Propane Decarbonizing effects the separation of asphalt by a difference in solubilities and at temperatures below 200° F. The asphalt separation is sharper than can be secured with vacuum distillation, and the fact that it is achieved at low temperature means that no thermal cracking occurs; thus the decarbonized oil is truly a virgin stock.

The process is distinguished by its high yields of clean catalytic-cracking stock; its ability to satisfactorily handle heavy, high-asphalt crudes which most refiners reject because of processing difficulties; and the fact that it concentrates any salt in the charging stock in the asphalt phase.

The asphalt produced as a by-product is a premium material for use in making finished asphalts for road-building, roof-surfacing, etc.

The propane lost in the process represents only about 0.1 percent of the propane circulated per day, and because of its low cost, is an unimportant factor in over-all operating cost. The total operating cost for decarbonizing catalytic-cracking feed stocks varies from four to nine cents per barrel of reduced crude, depending on the size of the unit and the cost of utilities.

A Propane Decarbonizing unit consists of a storage tank for liquid propane, charging equipment for propane and reduced crude, a countercurrent tower in which the actual asphalt separation is effected, and equipment for recovering propane.

The reduced crude is charged somewhat above the mid-point of the tower and the propane at the bottom of the tower. The majority of the asphalt is precipitated in the vicinity of the feed point. Since it is the heavier phase it flows over suitable baffles to the bottom of the tower. As the asphalt phase descends it is stripped of oil by upcoming propane with the result that only a high-melting-point, relatively oil-free, asphalt-propane solution reaches the base of the tower. This asphalt phase is continuously withdrawn, under flow control, from the bottom of the tower and proceeds to propane-recovery equipment. The oil-propane solution is further stripped of asphalt above the oil-feed point and passes from the top of the tower to its propane-recovery system.

Temperatures in the Decarbonizing Tower vary from about 120° F. to 180° F., depending on the character of the reduced crude. The heavier the reduced crude the lower the temperature at which a satisfactory separation is secured.

The ratio of propane to oil by volume normally varies from a low of about three volumes of propane to one of oil, to a high of five volumes of propane to one of oil. The lower ratios usually are suitable for heavy reduced crudes containing a high percentage of asphalt, whereas the lighter reduced crudes containing a smaller quantity of asphalt require a higher propane ratio.

The majority of the propane in both the decarbonized oil and asphalt phases is recovered under high pressure and sent directly to the propane condenser without any compression being required. The small amount of propane recovered in the final strippers by the use of open steam is first sent to a jet condenser where the steam is condensed, and the propane is then compressed to condenser pressure.

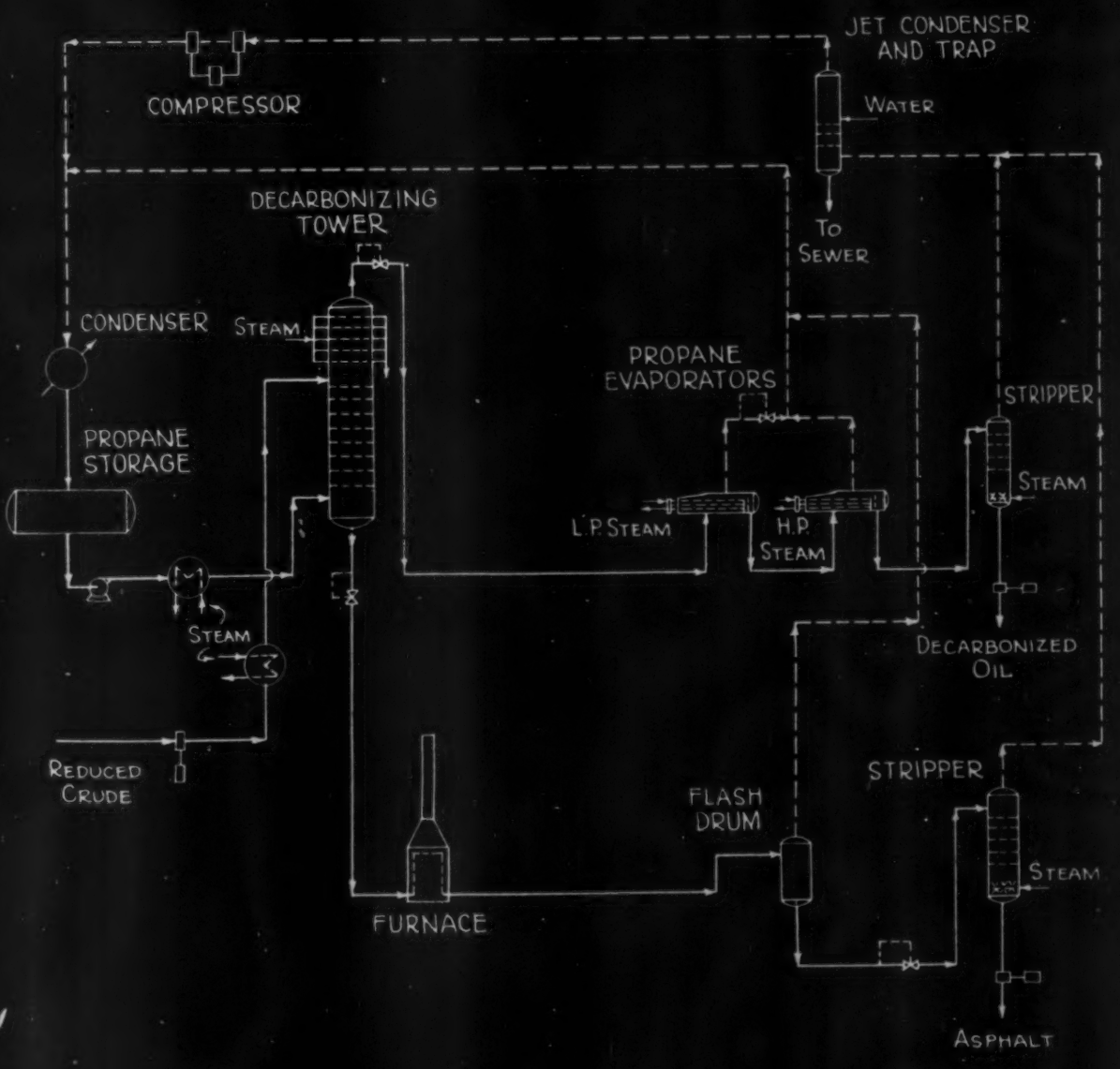
Typical operating results are found in the following table.

Table of Typical Operating Results—Propane Decarbonizing

REDUCED CRUDE	West Texas	Eastern Venezuela	Mississippi	South American
Vol Percent of Crude	20.0	27.6	57.0	61.0
°API Gravity	9.1	17.0	21.3	8.6
SSU Vis. at 210 °F.	.....	195	78.5	482*
Carbon Residue, Percent	18.7	6.7	7.9	.....
Flash Point COC °F.	645	430	275	.....
Fire Point COC °F.	730	525	345	.....
Decarbonized Oil:				
Vol. Percent of Crude	8.1	22.8	49.7	30.0
Vol. Percent of Red. Crude	40.5	82.5	87.2	49.2
°API Gravity	20.4	20.6	26.0	17.5
SSU Vis. at 210 °F.	305	99.1	49.4	224
Carbon Residue, Percent	3.4	2.2	1.2	2.6
Flash Point COC °F.	625	445	.....	500
Fire Point COC °F.	710	510	.....	575
Asphalt:				
Vol. Percent of Crude	11.9	4.8	7.3	31.5
Vol. Percent of Red. Crude	59.5	17.5	12.8	50.8
Specific Gravity	1.06	1.05	1.04	1.07
Melting Point, °F.	160	156	184	187

* Saybolt Furol Viscosity

The M. W. Kellogg Company has been appointed licensing agent for the process for the United States and foreign countries. Licenses granted are under the patents and patent rights of Standard Oil Company (New Jersey), Union Oil Company of California, Standard Oil Company (Indiana) and The M. W. Kellogg Company.



PROPANE DECARBONIZING PROCESS

(Tower Type)

COURTESY, THE M. W. KELLOGG COMPANY

BUILDERS FOR INDUSTRY
REFINERIES · CHEMICAL PLANTS · PIPE LINES
POWER PLANTS · PROCESS AND INDUSTRIAL PLANTS
Design · Engineering · Construction
BECHTEL CORPORATION
SAN FRANCISCO · LOS ANGELES
NEW YORK
BECHTEL

THIS NEW EQUIPMENT
REVOLUTIONIZES
SOLVENT EXTRACTION

Fractional Liquid Extraction Now Compares Favorably With Fractional Distillation

This equipment* should be considered for separations involving:

- Close-boiling mixtures.
- Azeotropic mixtures.
- Heat sensitive mixtures.
- Non-volatile mixtures.
- Materials from dilute solutions.
- Impurities and color-bodies.
- Isomers.

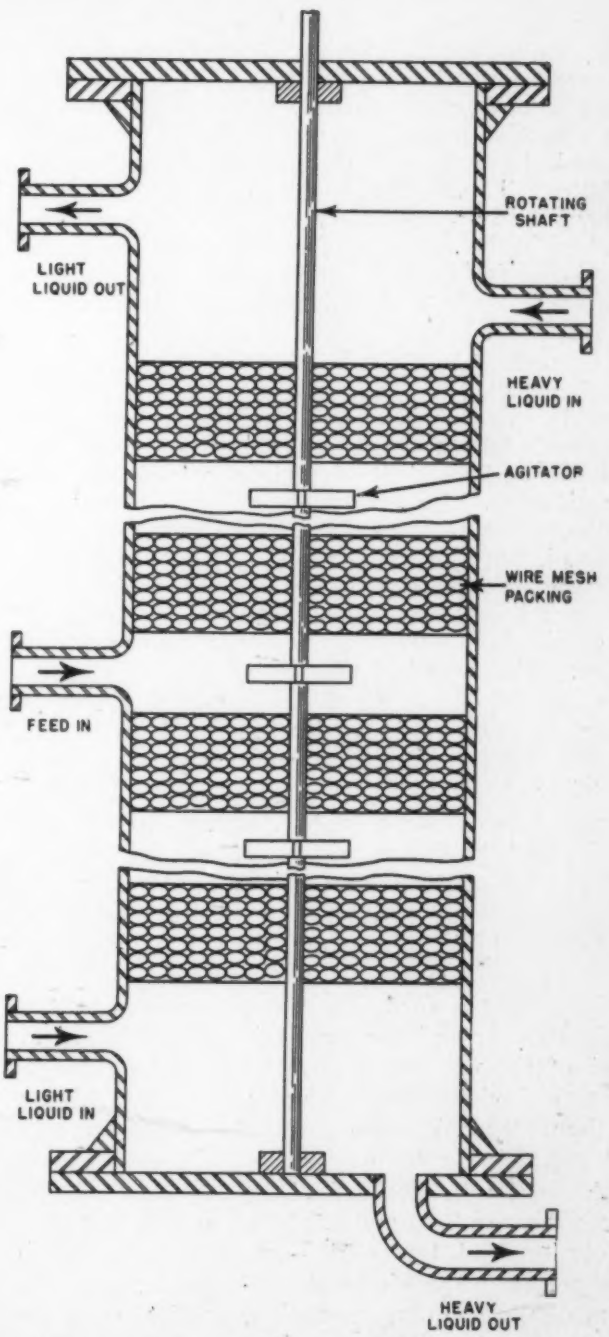
Unusual features of this equipment include:

- Low height per actual extraction stage.
- Any number of stages, as the process demands.
- 40 or 50 stages per column, if required.
- Low cost per stage.
- High throughput capacity.
- Excellent large scale performance.

Laboratory Unit

A 1" diameter unit for testing proposed plant scale separations and for general research work is available from stock. This extractor is a complete unit, including all the necessary auxiliary apparatus such as flow meters and control equipment.

YORK Mist and Entrainment Separators are used in atmospheric and vacuum distillation towers, gas absorbers, knock-out drums, separators and other process equipment to increase yields, to increase product purity, and to increase production capacity.



York
ENGINEERS AND MANUFACTURERS
* Pat. Applied for

Otto H. York Co., Inc.

360 Glenwood Avenue
East Orange, New Jersey
ORange 5-0316

Fractional Liquid Extraction Process

OTTO H. YORK COMPANY, INC.

FRACTIONAL liquid extraction is a process which is ideally suited to the separation of certain compounds which cannot be distilled or separated by ordinary distillation. The process depends on the different solubility properties of the components. These are a function of their molecular structure and are completely independent of their boiling points. Thus, it is possible to separate all the aromatic compounds from the paraffinic and naphthenic compounds, or all the paraffinic compounds from the naphthenic and aromatic. The first type of operation is required in the recovery of toluene from the crude oil fraction or from a hydoformer effluent. The second type of operation is used in the refining of lubricating oils.

A new type of liquid extraction column developed by E. G. Scheibel of Hoffman-LaRoche, Inc., makes it possible to obtain an equilibrium stage in liquid extraction in about the same column height as required for a theoretical plate in conventional distilling columns. The advantage of liquid extraction over distillation in these separations has long been recognized, but the application of the process has heretofore been restricted, because the problem of obtaining a liquid extraction column with twenty or more equilibrium stages in a reasonable size column at a reasonable cost had not been solved on a commercial scale unit.

The recovery of toluene by fractional liquid extraction is illustrated in the flowsheet on the opposite page. The toluene-rich fraction is separated in the fractional liquid extractor by the counter-current flow of a high boiling paraffinic solvent and phenol. In the section of the column above the feed the decending stream of phenol extracts the aromatics from the paraffinic solvent and the solution leaving the top of the column contains the paraffins and naphthenes in the feed. These are separated in the paraffin stripper and the paraffinic solvent from the bottom of the column is cooled by heat exchange against the feed and then with water to about 120°F. It is then recycled to the bottom of the fractional liquid extraction column.

In the section below the feed the paraffinic solvent extracts the paraffins and naphthenes from the phenol solution and leaves the aromatic compounds. These are recovered from the solution leaving the bottom of the column, by distillation in the phenol stripper. The phenol from the bottom of the column is partially cooled by heat exchange against the feed. It is finally cooled to about 120°F. and introduced into the top of the extractor.

The aromatic fraction from the overhead of the phenol stripper is run into the benzene tower to remove the benzene in the original feed. The bottoms from this tower is run into the toluene tower, where the toluene is taken overhead, and the xylenes and higher boiling aromatic compounds in the feed are removed from the bottom of the tower.

If the benzene and xylene fractions are not desired the total original feed can be fractionated to give a toluene heart-cut. The toluene can be sparated in the fractional liquid extractor to give a substantially pure toluene product from the phenol stripper.

Suitable modification of this process provides a practical method for the production of high cetane diesel fuel and high-octane aromatic blending stock.

The process can be applied also to the refining of lubricating oils by the use of solvents, which are more volatile than the oil fractions. In this case the paraffinic solvent would consist of a more volatile mixture, such as hexane and heptane, and the second solvent could be any one of those usually used in the refining of lubricating oils, such as nitrobenzene or phenol.

The fractional liquid extraction process also has applications in the separation of the water soluble by-products of the Fischer-Tropsch synthesis for liquid fuel. The process allows a separation of the acids, the alcohols, and the ketones aldehydes into three individual groups from which the components can be recovered in a pure state by ordinary distillation. It can also be used to overcome any difficulties encountered in the distillation of these groups such as in the separation of isopropyl alcohol, ethyl alcohol, and n-propyl alcohol, and the four isomeric butyl alcohols. Since water can be used as one of the solvents in these extractions, it is unnecessary to dehydrate the mixture of water soluble compounds to obtain a separation. This is a relatively difficult operation because they have a fairly wide boiling range above and below the boiling point of water and most of them form constant boiling mixtures with water.

Another application of the liquid extraction equipment and the fractional extraction process is in the recovery of the oil soluble oxygenated by-products from the Fischer-Tropsch process and their separation into useful products. This subject has not yet been publicized very widely, but it will be an important essential source for such materials when the synthetic liquid fuel plants are completed.

The fractional extraction process here outlined is economically practical for both small plants and for large scale operations. Where small quantities of valuable products are to be recovered, the equipment can be built to handle quantities expressed in pounds per day. For petroleum products and large tonnage chemicals, the capacity ratings are expressed in barrels or tons per day. The process lends itself readily to automatic control.

Additional information on this fractional liquid extraction process and on equipment required for the process is available from Otto H. York Company, Inc., East Orange, New Jersey.

FRACTIONAL EXTRACTION PROCESS
COURTESY OTTO H. YORK COMPANY, INC.
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
TOLUENE-RICH FRACTION
1000
FRACTIONAL LIQUID EXTRACTOR
120
120
FEED PREHEATER
6600
6000
PARAFFIN STRIPPER
250
180
0
100
NON-AROMATIC FRACTION
600
800
450
PHENOL STRIPPER
250
180
0
80
5400
5000
120
380
AROMATIC FRACTION
400
BENZENE TOWER
180
260
0
100
BENZENE PRODUCT
100
400
300
TOLUENE TOWER
240
220
0
400
TOLUENE PRODUCT
100
200
300
100
100
XYLENE PRODUCT

POSITIVE, SAFE AND AUTOMATIC

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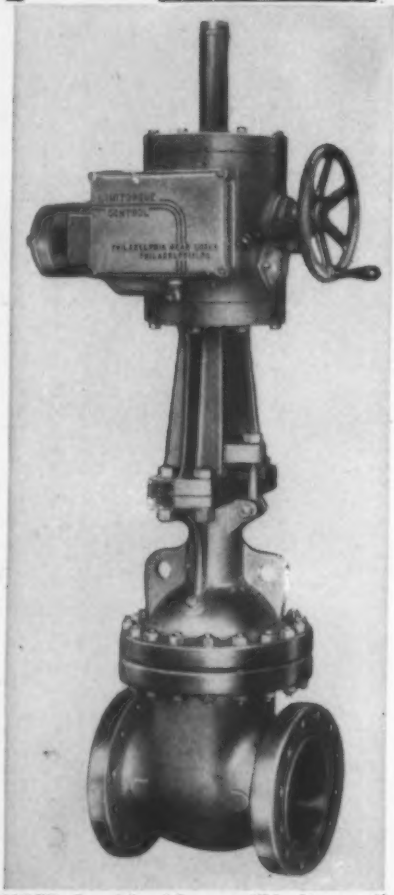

NOTE: Send for 90 page "Limitorque" Catalog—and please use your Business Letterhead when requesting same

-has been installed in the following Petroleum Refineries and Pipe Lines

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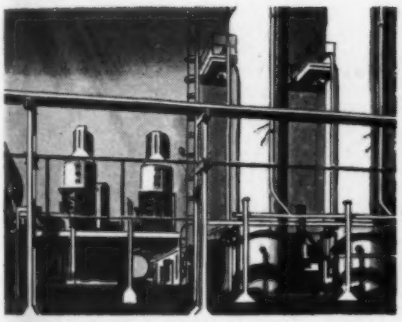
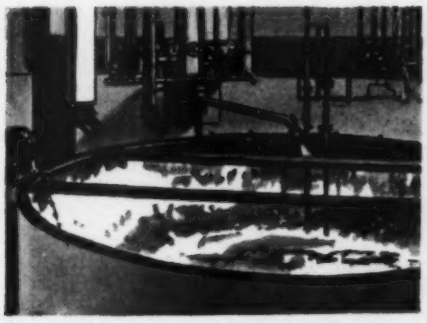
Are Your Corrosion Problems As Severe As These?

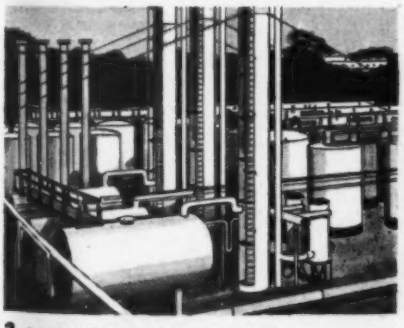
You won't be discouraged by any corrosion problem, when you read how these four companies have finally controlled extremely severe corrosion in their plants with a system of UCILON Protective Coatings!



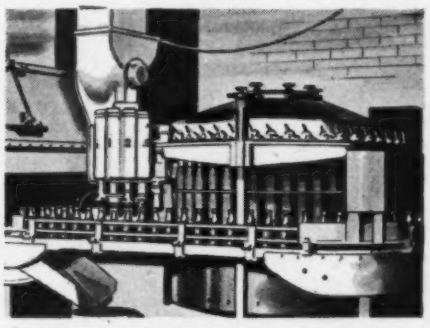
1. A WELL KNOWN REFINERY had Ucilon coatings applied to ducts exhausting oil and gasoline fumes at the cracking tower. One year after application, there was still not a sign of failure, despite the severity of the service.



2. A SOAP PLANT had a tough ceiling maintenance problem—some ordinary paints being destroyed by fatty acid vapors in as little as 2 weeks. Yet after *6 months* of exposure, Ucilon coatings were still totally unaffected!



3. A CHEMICAL PROCESSOR coated a steel storage tank with a Ucilon coating system. Especially hard on protective coatings, the tank's contents broke down two other coatings tried —one in 6 weeks. Yet Ucilon is still in good condition after *14 months!*



4. IN THE BOTTLING ROOM of one company, painting of machines was considered to be almost useless—until they discovered Ucilon. Nothing ever tried even approaches the effectiveness of Ucilon, which for *2 years* has been withstanding the humidity and spillage!

Similar reports continue to arrive from plants of every nature. Maintenance men are discovering a degree of corrosion protection and service life they never achieved with any coating tried before. If *you* have an exceptionally tough coating problem, we're confident that there's a Ucilon coating system to help you stop corrosion damage and vastly lengthen the period between your paint jobs. Let us send you our specific recommendations without obligation. Write today, outlining the details of your application.

UCILON* 400

Ucilon 400 is a flexible, glossy, air-drying coating for metal, concrete and wood. It is applied by brushing or spraying. When solvents evaporate, a tough "plastic" film is formed. Ucilon 400 provides exceptional resistance to acids, alkalies, alcohols, oils and greases, salts, water, cleaning compounds, oxidizing agents.

OTHER UCILON COATINGS

Ucilon 1601—a ready-mixed, extra-bright aluminum paint with exceptional outdoor resistance. It stays bright under prolonged exposure, withstands elevated temperatures.

•

Ucilon 1501—a very hard, chip-resistant baking coating which resists chemical attack at temperatures higher than recommended for Ucilon 400.

SEND FOR THIS BULLETIN *which gives more details on various Ucilon coatings and their properties.*

*Trade Mark Reg. U. S. Pat. Off.

UCILON Protective Coatings

UNICHROME products of UNITED CHROMIUM, INCORPORATED

51 East 42nd St., New York 17, N.Y. • Detroit 7, Mich. • Waterbury 90, Conn. • Chicago 4, Ill. • Dayton 2, Ohio • Los Angeles 13, Cal.

Solvent Dewaxing Process

TEXACO DEVELOPMENT CORPORATION

THE Solvent Dewaxing Process generally employs a mixture of dimethyl ketone (acetone) or methyl-ethyl-ketone and benzol as the solvent to obtain a dewaxed-oil pour point equal to or within a few degrees of the dawaxing temperature. The dewaxed oils are free from wax cloud and there is no rise in pour point or "pour reversion" on aging. Either distillate or residual stocks in the raw state or refined, of any viscosity, can be dewaxed. Texaco Development Corporation, 135 East 42nd Street, New York, is licensor of the process.

The process is widely used in both the United States and Europe. Thirty-six units have been installed, with a total daily charging capacity of over 72,000 barrels, processing a wide variety of lubricating-oil stocks of high and low wax content. Contracts have been awarded for 20 additional units having a total charging capacity of over 55,000 barrels per day.

Simplicity of operation, flexibility, low construction and operation costs, the absence of special equipment, special alloys, high pressure, excessively low operating temperatures and the ability to obtain high dewaxed-oil yields and very low oil-content waxes are outstanding characteristics of the process.

The wax contained in lubricating oil stocks when properly precipitated from a chilled solution in the solvent mix is particularly adaptable to removal by filtration. Practically complete separation of oil from wax is obtained by use of a cold displacement wash. The use of a closed flue-gas system minimizes fire hazards and solvent losses. Dilution solvent dosages vary from a solvent-to-oil ratio of 1.0:1 to as much as 4:1, depending upon the nature and viscosity of the charge stock.

Recently improved design of exchange and chilling equipment has resulted in reduced refrigeration requirement. Proper conditioning of the charge to the filters by controlled application of solvent has increased dewaxed-oil yield, reduced oil content of the wax to an extremely low figure, and reduced the amount of solvent circulated per barrel of oil charged. An additional development has been the production of wax, without addition agents, concurrently with the production of low-pour-point oil, but in auxiliary equipment operated as a separate process. White crystalline waxes and higher-melting-point microcrystalline waxes are produced from distillate and residual stocks, respectively, to meet specific requirements.

The accompanying flow diagram, upon which it is impractical to reflect all of the recent improvements, shows the flow through the principal chilling, filtering and solvent-recovery equipment. Dilution solvent is introduced into the charge-oil stream in such amounts, at selected points in the chilling cycle, as to insure a wax crystal structure and liquid viscosity most suitable for filtration. The chilled charge mixture flows from the double-pipe chillers through a filter-feed tank to drum-type vacuum filters in which compartmentized, cloth-covered drums rotate half submerged in enclosed filter cases. A wax-free filtrate solution is drawn through the filter cloth to filtrate tanks in which the vacuum which induces filtration is maintained. The wax cake deposited upon the drum during filtration is washed with cold solvent as it emerges from the liquid level in the filter; much of the wash solvent being drawn into the filtrate system toward the end of the filtration cycle. Just before each of the filter-drum compartments reaches the wax deflector, flue gas at low pressure is admitted through a blow-back connection to the underneath side of the filter cloth. The bond between wax and cloth is broken and the washed wax falls into the wax-discharge compartment. Controllers in the flue gas piping maintain atmospheric pressure in the filter hood.

The filtrate valve on the drum trunnion is so constructed that oil-rich filtrate collected during the early portion of the filter cycle is directed to the main filtrate tank, whereas the oil-lean filtrate collected late in the cycle is directed to the wash filtrate tank. The oil-lean wash filrate then may be returned to the charge-oil stream as a substitute for a portion of the dilution solvent. Occasional warm washing of the filter cloth is a simple operation requiring 15 to 20 minutes and is conducted, without interruption to uniform flow through the plant. The warm washings, consisting primarily of solvent, generally are returned to the charge mixture, this further reducing the requirements for new solvent.

The filtrate from the main filtrate receiver is pumped through the double-pipe exchangers to evaporators for the recovery of the solvent. Heat for evaporating the solvent from the dewaxed-oil solution and wax mix may be supplied by either steam or fired heaters.

Substantially all of the water entering the system is crystallized and removed with the wax to the wax-mix surge and settling tank. This vessel is maintained at about 130° F. under which condition melted ice settles to the bottom of the tank. To this water is added condensates from steam strippers and the water-bearing material is pumped to a ketone fractionator for recovery of solvent and removal of water from the system.

The process produces high yields of low-pour-test oils free from wax cloud. Due to low solubility of wax in the solvent, low-pour tests can be obtained at relatively high separation temperatures. Extremely low pour points can be obtained, due to the narrow difference between the separation temperature and the resultant pour point.

Consistent uniformity in operating conditions and product quality is an inherent characteristic of the process. There are no corrosion or erosion problems and low-pressure equipment is used throughout.

SOLVENT DEWAXING PROCESS
COURTESY, TEXACO DEVELOPMENT CORPORATION
KETONE FRACTIONATOR
KETONE CONDENSER
STEAM & SOLVENT CONDENSER
WAX-FREE OIL STORAGE
WAX STORAGE
WAX MIX SURGE & WATER SETTLING TANK
WAX MIX HEATER
WAX STEAM STRIPPER
WAX TO WAX MIX EXCHANGER
WAX MIX FLASH SECTION
DEWAXED OIL STEAM STRIPPER
FINISHED OIL COOLER
FLUE GAS CHILLER
FLUE GAS AFTER COOLER
VAC PUMP
FILTRATE VAC. RECEIVER
CHILLER
EXCHANGER
WAX MIX FLOW TANK
WAX CONVEYOR
SOLVENT STORAGE
SOLVENT CONDENSER
DEWAXED OIL MIX L.P. FLASH SECTION
DEWAXED OIL MIX H.P. FLASH SECTION
DRY SOLVENT RECEIVER
DEWAXED OIL MIX HEATER
CONTINUOUS VACUUM FILTER
VAPOR HEAT EXCHANGER
COOLER
SOLVENT RECEIVER
FILTER FEED FLOW TANK
CHILLER
EXCHANGER
COOLER
HEATER
DEWAXED OIL MIX SURGE TANK
WAX OIL CHARGE
LEGEND

"Hold It Joe—This Is A Homestead Lever-Seald"

Homestead
LEVER-SEALD
Valves

ALWAYS WORK
LAST LONGER
COST LESS PER YEAR

STICKING valves are never a problem with Homestead Lever-Sealds on the job. Built into each valve is a powerful leverage, ready at all times to overcome the effects of corrosion, viscous fluids or temperature and pressure extremes which cause ordinary valves to stick or "freeze."

This powerful leverage not only relieves pressure between the seating surfaces to permit operation under all conditions; but it also is the means of providing the extra tight seal against leakage—the seal which keeps corrosive or erosive line fluids away from the seating surfaces; assures longer valve life and lowest cost per year.

Made in stainless steel, steel, semi-steel, bronze, monel or other alloys suitable for the working conditions under which they are to be used, Homestead Lever-Seald Valves may be had in sizes from 1" to 12"; and for working pressures from 150 pounds to 1500 pounds.

Choose the sizes and types for your services from Reference Book 39 Section 3. Write for your free copy today.

Check these Outstanding Features:

A QUARTER-TURN fully opens or closes valve.

INSTANT STICK-PROOF OPERATION. Plug turns freely after seating pressure is relieved by screw lever.

POSITIVE LEAKLESS SEAL is obtained by applying full seating pressure with screw lever after opening or closing.

LEAK PROOF. Deep stuffing box, gland and sealed bottom prevent fluid loss, and reduce hazards.

STRAIGHT-LINE FLUID FLOW. Little or no pressure drop.

SEATING SURFACES ALWAYS PROTECTED in both open and closed positions. Fluid or grit cannot damage valve seat.

HOMESTEAD VALVE MANUFACTURING CO.

P. O. BOX 16 CORAOPOLIS, PENNA.

SPARKLER MANUFACTURING COMPANY

296 Lake Street, MUNDELEIN, ILL.

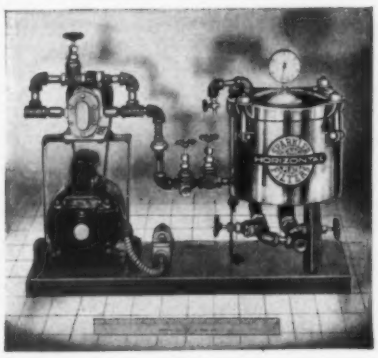
Manufacturers of High Grade Industrial Pressure Filters

Units for Filtering Every Type of Petroleum Product

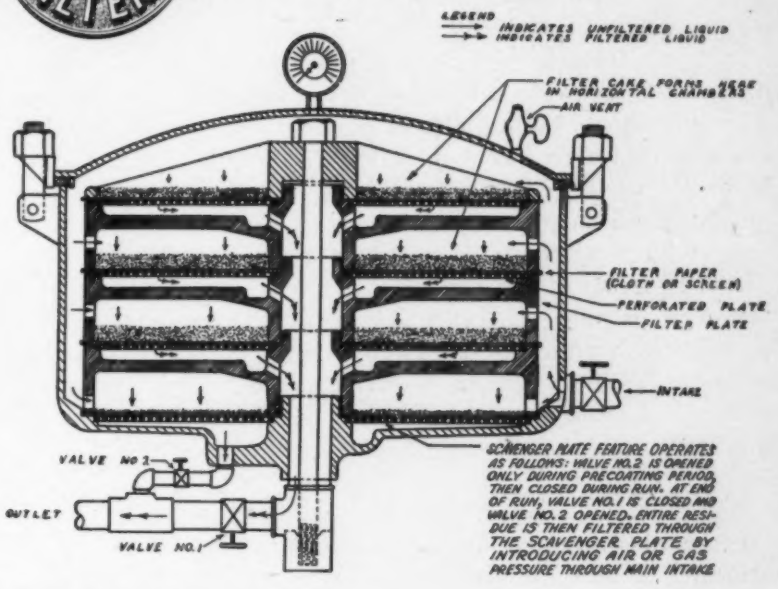
Gasoline, Distillates, Catalyzers, Oils, Lubricants, Solvents, Paraffins, Waxes, Alcohols, Chemicals, Etc.

SPARKLER HORIZONTAL PLATE FILTERS

Engineering "know-how," plus many years of experience in setting up successful installations, is the reason so many modern refineries are solving their filtration problems with Sparkler HORIZONTAL PLATE Filters.



Standard Model 8-3 Pilot Plant Filter in stainless steel—also made in carbon steel.


LEGEND
INDICATES UNFILTERED LIQUID
INDICATES FILTERED LIQUID
FILTER CAKE FORMS HERE ON HORIZONTAL CHAMBERS
AIR VENT
FILTER PAPER (CLOTH OR SCREEN)
PERFORATED PLATE
FILTER PLATE
INTAKE
VALVE NO 2
OUTLET
VALVE NO. 1
SCAVENGER PLATE FEATURE OPERATES AS FOLLOWS: VALVE NO. 2 IS OPENED ONLY DURING PRECOATING PERIOD, THEN CLOSED DURING RUN. AT END OF RUN, VALVE NO. 1 IS CLOSED AND VALVE NO. 2 OPENED. ENTIRE RESIDUE IS THEN FILTERED THROUGH THE SCAVENGER PLATE BY INTRODUCING AIR OR GAS PRESSURE THROUGH MAIN INTAKE

SPARKLER FILTERS give you —

- uniform high quality filtrate to end of cycle
- stable filter cake
- complete batch filtration
- flexibility
  continuous or intermittent
  portable or stationary
- high flow rates
- long cycles
- easy cleaning
- no leakage
- low cost operation

ENGINEERING DATA

Pumps: Standard models equipped with either rotary or centrifugal pumps. Horsepower of motor depends on the specific gravity and viscosity of filtrate and the pressure range necessary.

Filter tanks: Welded construction in iron, bronze, rubber-lined mild steel, nickel, monel, all types of stainless steel, or inconel.

Filter plates: Spun and welded construction in above materials; also in Haveg and molded hard rubber. Made in 8¼, 13½, 18 and 33 inch diameters. Perforated screens to fit made in above metals, molded hard rubber, and plastics. Plates in shallow (S) or deep (D) construction (see model numbers), thus accurately proportioning cake space to filtering area for specific application. (Pressure drop across plates is a direct function of viscosity of filtrate and thickness of cake.)

Mountings: Filters for in-line installation mounted on pipe legs with floor flanges. Complete filters are mounted with pump and motor on mild steel stand with ball-bearing casters. Laboratory models are mounted on steel base with rubber bumpers.

Jackets: Steam or brine jackets in mild steel built for 15 lb. p.s.i. working pressure. Jackets for higher pressures on special order only.

SOME REPRESENTATIVE SIZES

MODEL No.	Capacity G.P.H.	Area Sq. Ft.	Cake Space Cu. Ft.
8-3	30–60	1.0	144 Cu. In.
18-D-8	650	13.57	2.0
33-D-12	3500	70.0	11.7
33-S-17	5000	99.48	8.3
33-S-26	7600	152.0	12.65

MANY MODELS AND SIZES: Capacities 60 to 10,000 G.P.H.
Free Filtration survey on request.

Model 33-S-9 in carbon steel. Steam jacketed. Cap. 2,600 G.P.H.

Propane Dewaxing Process

THE M. W. KELLOGG COMPANY

PROPANE dewaxing is a well-known lubricating-oil process for separating wax from crude oil. The solvent, propane, is the only extraneous material used in the process since it serves both as a refrigerant and as a solvent for the oil.

Economic features of the process are its low operating cost and wide adaptability to any type of lubricating stock. High yields of dewaxed oil of zero pour test or lower and wax of low oil content are obtained.

Eight plants now operating or being built in this country have a yearly capacity of approximately 7,500,000 barrels of charging stock.

Oil and propane are charged continuously under flow control, warmed to dissolve all of the wax, and sent to a chilling vessel in which reduction of temperature to approximately —40° F. is effected by allowing propane to vaporize directly from the solution.

Wax crystallizes out at this low temperature and is separated from the oil on a continuous rotary pressure filter of most modern design.

The filter cake is washed with cold liquid propane to remove oil adhering to the wax crystals and then transferred by means of a screw conveyor and wax pump to propane-recovery equipment.

The dewaxed oil solution also is sent to a propane-recovery unit. Both end products—dewaxed oil and wax—are pumped from the strippers to storage tanks.

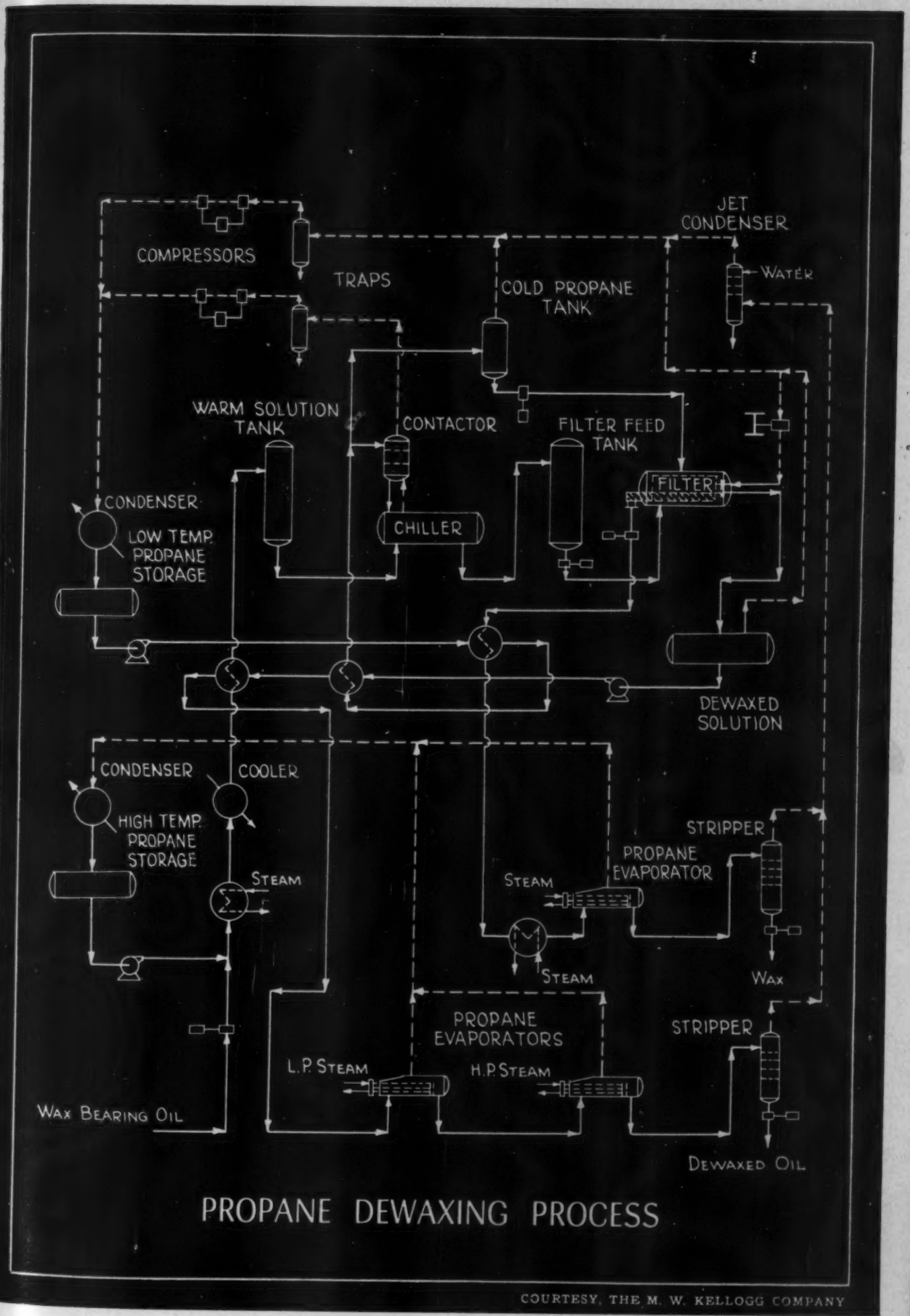
The low cost of propane, its double-duty as a solvent and refrigerant, and its ease of recovery are features of the process which make for high efficiency.

Propane dewaxing often is employed in combination with propane deasphalting and acid treatment in propane solution. The following table gives pertinent data on propane dewaxing of four lubricating stocks:

Propane Dewaxing Process

	Propane Deasph. Phenol Treated 15% East Texas Reduced Crude	Phenol Treated East Texas Heavy Distillate	Propane Deasph. Phenol Treated West Edmonds SAE 70 Stock	Propane Deasph. Phenol Treated Oklahoma City SAE 40 Stock
Propane, Volume Percent	300	250	300	300
Filtering Temperature, °F.	—34	—35	—40	—41
Dewaxed Oil, Vol. Percent	61.6	67.7	85.5	81.1
Inspections Waxy Oil—				
°API Gravity	29.7	32.0	27.4	30.1
SSU Vis. at 210 °F.	93.1	57.3	132	72.7
Color—NPA	4½—	2—	3½—	2—
Carbon Residue, Percent	0.20	....	0.25	0.06
V.G.C.	.795	.790	.803	.797
Dewaxed Oil:—				
°API Gravity	27.6	28.9	26.7	28.7
SSU Vis. at 210 °F.	118.4	66.2	150.0	79.6
Color—NPA	4½—	2—	3½—	2—
Pour Point, °F.	0	0	0	—5
Carbon Residue, Percent	0.26	0.03	0.30	.05
Viscosity Index	95	92.1	91.1	92.1
V.G.C.	.804	.808	.807	.804

The M. W. Kellogg Company has been appointed licensing agent for the process for the United States and foreign countries. Licenses granted are under the patents and patent rights of Standard Oil Company (New Jersey), Union Oil Company of California, Standard Oil Company (Indiana) and The M. W. Kellogg Company.



PROPANE DEWAXING PROCESS

COURTESY, THE M. W. KELLOGG COMPANY

INTERNATIONAL ENGINEERING, INC.

DAYTON 1, OHIO, U.S.A.

15 PARK ROW, NEW YORK CITY 7 — **407 S. DEARBORN, CHICAGO 5**

GASOLINE BLENDING AND MIXING EQUIPMENT

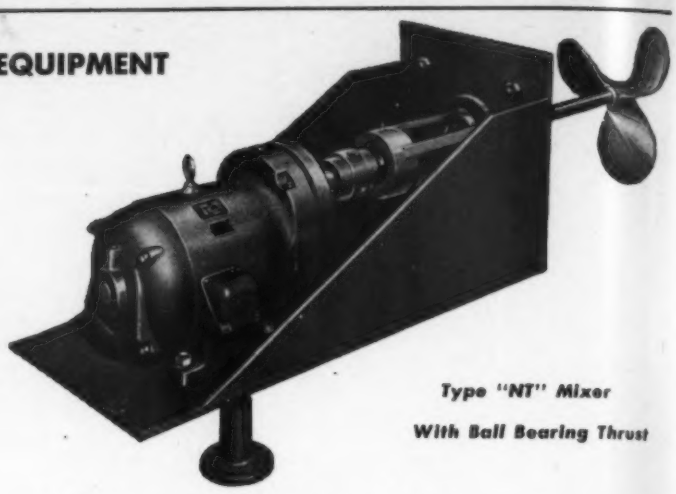
INTERNATIONAL SIDE ENTRANCE MIXERS, due to their high efficiency, adapt themselves most readily to the mixing of large batches of gasoline with tetraethyl. They blend motor fuels with a time saving of from 70% to 95%—release tanks for charging in shorter time—save much power—increase production without installation of additional tanks—fitted to any type of large tank, whether spheroid, A.P.I. rafter roof or pontoon—decrease time of tank car or tanker hold-over—cut evaporation losses—cut maintenance cost—save time and expense in caustic washing operations, lubrication oil blending, etc.

INTERNATIONAL SIDE ENTRANCE MIXERS, Type "N," consist of a strong fabricated plate base which base contains a Master Electric or equivalent ball bearing, explosion proof gear motor connected by means of a flexible coupling to the mixer shaft.

The mixer shaft operates in a double lantern bearing stuffing box (Type "B" double bearing) which stuffing box is the latest design now offered to the refining industry. The shafts are over-large and the propellers are machined to size, 3-blade Marine type, thus being more efficient than the ordinary cast, snag wheel ground and polished propellers. Mixer is also fitted with an adjustable pipe support for alignment.

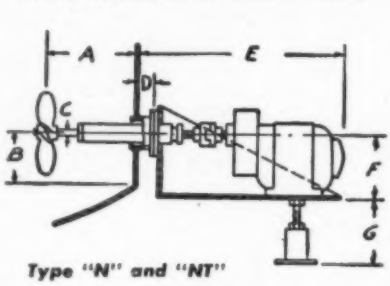
INTERNATIONAL TYPE "NS" MIXERS are a duplicate of the Type "N" excepting that they use a V-belt and separate motor drive instead of a gear motor.

Mixers are mounted on nozzles as shown or manhole covers and can be arranged with special connections for wood or other special tanks.

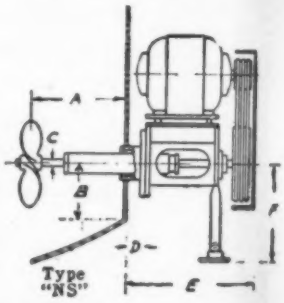


Type "NT" Mixer With Ball Bearing Thrust

TYPE "NT" MIXERS are the most improved design Side-Entrance Mixers offered to the trade. They are fitted with a double lantern bearing stuffing box and ball bearing thrust, easily repacked from the outside.



Type "N" and "NT"



Type "NS"

Type "N" Easily Repacked from Outside.

Ask for New Catalog No. 115

SIDE ENTRANCE—SPECIFICATIONS

Propeller Dia. Variable with Liquid

Type No. "N" or "NT"	Motor H.P.	Standard Shaft Speed	DIMENSIONS Type "N" and "NT" Stuffing Box "B" A	B	C	D	E-"N"	E-"NT"	F	G
N-1	½	....	8	6	1¼	4	31	35	7⅞	7
N-2	¾	400	8	6	1¼	4	31	35	7⅞	7
N-3	1	400	8½	7	1¼	4	32	36	7⅞	7
N-4	1½	400	9	8	1¼	4	33	37	8	9
N-5	2	400	10	9	1¼	4	38	43	8⅜	8
N-6	3	400	12	10	1¼	4	38	43	8⅜	8
N-7	5	400	14	11	1½	4	43	44	10⅜	8
N-8	7½	400	16	12	1½	4	42	48	12¼	9
N-9	10	400	18	13	1½	4	46	53	14¾	9
N-10	15	400	24	14	1¾	4	48	56	14¾	9
N-11	20	400	30	15	2	4	53	62	16½	9
N-12	25	400	36	16	2½	4	56	67	16½	9

TYPE "NS"—V-BELT

NS-1	½	400	8	6	1	4	19½	....	14	..
NS-2	¾	400 Standard 1150 Maximum	8	6	1	4	20½	....	14	..
NS-3	1		8½	7	1¼	4	21	....	14	..
NS-4	1½		9	8	1¼	4	22½	....	14	..
NS-5	2		10	9	1¼	4	23	....	16	..
NS-6	3		12	10	1¼	4	25½	....	16	..
NS-7	5		14	11	1½	4	28½	....	16	..

TOP ENTERING MIXERS

International manufactures a complete line of permanent nozzle style mixers, as shown in International catalog No. 110. These are the latest design Mixers offered the process industry, using either propellers or turbines.

PORTABLE MIXERS

International also manufactures a complete line of the latest in portable mixers, either electric or air-driven. These mixers are well fitted for the blending of special oil batches, suspending lime, blending of medicinal oils, agitation of slurries, as well as for excellent use in synthetic rubber pilot plants, etc.

Ask for Catalog No. 115

FLOWRATOR instruments

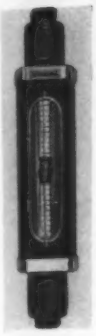
TRADEMARK

for every flow measurement

* FLOWRATOR is the trade mark name of Fischer & Porter Company to designate its instruments for flow measurement and automatic flow control. It is an "area-type" meter in which the variation in area is obtained by means of a vertical tapered tube with the smallest diameter at the bottom. A weight of fixed diameter (called "float") moves freely within this tapered tube, and hence the area between the float and tube increases as the float rises. This area then becomes a measure of the flow and for every flow rate there is a fixed position of the float in the tube.

Write for *Catalog 10* in which we illustrate in an easy-to-read style the many advantages of this method of flow measurement.

"MASTER-ENCLOSED" FLOWRATOR METER



Its important advantages are:

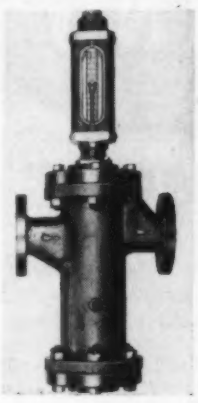
1. Resistance to corrosion—available in practically any material.
2. Easy to disassemble.
3. Flanged packing glands—adjustable from outside the meter.
4. Handsome appearance—ideal for "good-looking" panel mounting.

This versatile flow meter can be adapted to practically all the many diverse requirements of the various industries. Available with flanged or screwed fittings . . . with vertical or horizontal connections. Also in a "Four Post Open" design with tube and float visible from any angle at any distance or with heavy rugged side plate construction for protection against excessive pipe strain and external blows.

Send for *Catalog 25.*

ARMORED FLOWRATOR METER

For use in installations where high pressures, high temperatures, possibility of liquid hammer or the handling of hazardous fluids makes the use of the standard FLOWRATOR METER impracticable.

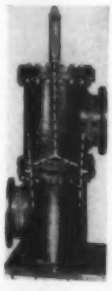


Chief Advantages:

1. The Armored FLOWRATOR METER may be piped horizontally and, therefore, easily replaces orifice or venturi meters for steam and other fluids.
2. Its one-piece body will take full pipe line strain.
3. It may be built readily to withstand high pressures, temperatures.

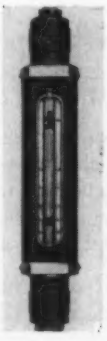
Standard materials of construction for the body are cast iron, steel, bronze, or stainless steels. For highly corrosive services, it may also be obtained in special materials such as aluminum, lead, monel, hastelloy, etc. For a complete description, dimensions and capacities, send for *Catalog 40.*

RATOSLEEVE FLOWRATOR METER



With the constant increase in the acceptance of the variable-area type flow meter as a superior type of flow rate instrument, there has been a steady trend toward handling larger and larger flows by means of it. In order to handle large flows with a reasonably priced meter, the new Ratosleeve has been devised. Essentially the Ratosleeve consists of a tapered metering sleeve insert that fits between two standard pipe tees in a manner similar to an orifice plate. It will handle up to 14,000 GPM of water or 16,000 CFM of air. Send for *Catalog 83.*

CONTINUOUS VISCOSIMETER



This VISCOSIMETER permits instantaneous observation of viscosity values existing in a moving fluid stream under full line pressure. It is available as a simple indicator or for continuously recording or automatically controlling viscosity values. This instrument is particularly valuable for continuously blending lube oils, maintaining constant fuel oil viscosity for improved oil burner operation or indicating end points in various plastic processes.

Send for *Catalog 88.*

RATOMATIC
Recorder-Controller

RATIO-CONTROLLER

REMOTE INDICATING, RECORDING, CONTROLLING AND PROPORTIONING INSTRUMENTS

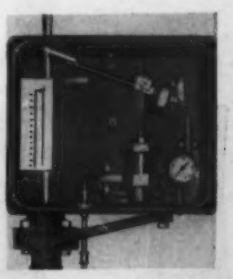
A complete line of secondary instruments has been developed by F & P Co. so that FLOWRATOR METERS may be provided with remote indicating, recording, totalizing, automatic controlling and automatic proportioning arrangements. These secondary instruments are connected to the primary FLOWRATOR METER through induction coil transmitters or magnetic transmitters. F & P has made a noteworthy series of advances in the field of inductive transmission which has brought the direct acting inductive transmitter—as against the null balance relay type—to new standards of reliability. Null balance instruments are, however, also supplied where highest accuracy is necessary despite higher cost.

When explosion hazards or other considerations require the use of secondary instruments using no electric current, the F & P "MAGNA-BOND" transmitter —a new type of powerful magnetic coupling—is used between primary FLOWRATOR METER and secondary instrument. The readings of this instrument may then be carried to remote locations by pneumatic transmitters.

Flow rates as low as 1 cc/min can be remotely indicated, recorded, controlled and totalized with the RATO-TRONIC which electronically amplifies FLOWRATOR METER float position.

Catalogs 50, 52, 53 and *55* available on request.

RATO-TRONIC
For Miniature Flow Rates



MAGNA-BOND
Pneumatic-Transmitter

FISCHER & PORTER CO., DEPT. IC-HB, HATBORO, PENNA.

PETRECO *Electric* DESALTING

DOES MORE than just remove salt...

One of the most important and profitable benefits refiners gain from Petreco Electric Desalting is *increased refining capacity*. This ability to charge more crude results directly from the high efficiency of Petreco Desalting, which

1. Eliminates salt plugging in condensers and tower trays.
2. Reduces corrosion from evolved HCl.
3. Prevents the formation of hard, flinty coke in furnaces.
4. Removes suspended solids such as sand, silt, rust, etc., from the crude charge.
5. Traps and removes slugs of water or B.S. picked up from crude storage.

By forestalling the basic difficulties inherent in plugging, corrosion, coking, foreign solids and B.S. and W. in the crude charge, Petreco Desalting

1. Increases heat exchange efficiency.
2. Reduces mechanical shut-downs caused by erosion and corrosion.
3. Permits high furnace temperatures, deeper cracking.
4. Keeps the refinery on stream longer, reduces down-time.
5. Cuts maintenance and repair costs.

Besides insuring efficient salt removal, PETRECO provides complete service, available replacement parts, engineering cooperation, and laboratory assistance, and the KNOW HOW gained from many years of desalting experience.

For further information, or a conference on your salt problem, just write or call any Petreco headquarters. A Petreco engineer will call on you, at your convenience. Get *all* the facts,–it pays.

PETROLEUM RECTIFYING COMPANY { 5121 S. Wayside Drive, Houston 1, Texas
648 Edison Building, Toledo 4, Ohio
530 W. Sixth Street, Los Angeles 14, Calif.

PETRECO

SPECIALIZED *Electric* PETROLEUM PROCESSES { DESALTING
DEHYDRATING

Desalting, Filtration, Acid Treating Processes

SOME of the essential process steps employed in refineries have a more or less accessory relationship to the basic conversion, separation, and distillation operations. Feed stocks must be made suitable for processing to best advantage and products must be finished to established standards of quality.

The desalting of crude oils before charging to distillation equipment is an accepted practice in efficient refinery operations, except in those increasingly rare instances where original salt contents are satisfactorily low. The economic necessity for obtaining high throughput capacities together with long on-stream operating periods and lower maintenance costs has led plant engineers to give careful attention to crude oil desalting. Desalting prevents the plugging of heater coils, greatly lessens equipment corrosion losses, and improves product quality. Several desalting methods are in general commercial use with excellent results, so that the choice of method is largely a matter of suitability to the specific crude oil and to plant processing factors.

Long established practice in process plants for the finish refining of lubricants, some distillates, and various specialty products involves the contacting of the raw product with an adsorbent and subsequent separation by filtration. The use of adsorbents is predicated on their ability to remove suspended, colloidal, and dissolved impurities, carbon or coke and resinous and asphaltic substances, from the oil treated. This includes hydrocarbons with high percentages of combined oxygen, those low in hydrogen content, and those containing coloring matter. Adsorbents are used also to neutralize acid-treated oils as an alternate to alkaline solution washing. Early practice in contacting oils with adsorbents, still followed with excellent results in many plants, is the gravity filtration of the oil through beds of the adsorbent contained in large special tanks. More recently, continuous contact and filtration processes have been devised which allow large volumes of oil to be treated in a plant of comparatively small ground area. Greater flexibility in adjusting operating conditions, particularly temperature, and also in changing between the products treated, is achieved. A considerable number of such plants are in operation.

Treatment with sulfuric acid has been used for the entire range of refinery products since the early days of the industry. Acid treatment has marked advantages for such treating operations and has undoubtedly been the most universally employed of all methods. More recently, the development of a variety of treating processes especially applicable to the improvement of gasolines and distillates and the widespread adoption of solvent refining methods in the processing of lubricants has tended to make acid treating less conspicuously the universal method. Even so, there are many instances where no substitute has been found which promises overall improvement when all factors of economics and operation for the specific case are considered, and acid treating procedures continue to occupy an important place in the process schemes of many refineries.

Crude Oil Desalting Process

VISCO PRODUCTS COMPANY, INC.

ONE serious problem confronting the refiner is that the majority of crude oils are produced with contaminating water either as an emulsion with the oil or in free state.

An emulsion is a system containing two immiscible liquid phases, one dispersed as globules in the other. Both oil-in-water and water-in-oil types exist. In the first, oil is the disperse and water the continuous phase. In the second, and predominant type, oil is the continuous phase.

Crude oil emulsions, to be acceptable for pipe line transportation often have a specified maximum of 2 percent basic sediment and water. In some instances 0.5 percent is specified. The treatment of emulsions should be performed with minimum practical delay, since standing and/or additional agitation frequently results in further stabilization. When dispersed water globules approach a diameter in the range of 10 microns the emulsion is quite stable and resists treatment, storage tank bottoms being a typical example.

In refinery operations emulsified crude oil charged to stills can drastically reduce throughput and cause costly corrosion problems. Salt deposits in exchangers and furnace tubes will reduce thermal efficiencies and eventually plug equipment, resulting in costly shut down periods. Cracking equipment is particularly susceptible to corrosion and the residual salts from an untreated crude may hydrolyze during the process to form hydrochloric acid and other acid bodies which readily attack iron.

The maximum salt content which can be tolerated by refineries varies somewhat with the design of the plant. In general, refiners consider a salt content of 20 pounds per 1,000 barrels of oil as the poorest acceptable requirement for crude charge stock. In many modern plants this figure may approach 5 to 10 pounds per 1,000 barrels, because it has been found that some salts in particular poison the catalyst used in catalytic cracking. It is not uncommon for refineries to achieve upwards of 95 percent salt removal. The average possibly ranges about 90 percent.

Emulsions vary widely, and their characteristics, both quantitative and qualitative, must be considered before the method of treating can be selected. Basically, three methods of treating have evolved over the years and currently are in wide use: mechanical, electrical and chemical. Many variations and combinations of these basic methods have been successfully used. The application of heat has been more or less universally acknowledged desirable, regardless of the method of treating. Mechanical methods include heating and settling, and centrifugal separation.

Heat lowers the viscosity and interfacial tension of an emulsified oil and, also, alters the colloidal properties. Heat is not singularly responsible for the separation of oil and water, but is an accelerator. If the difference in specific gravity of components is increased to the point where the water will precipitate, the separation or stratification is not due entirely to heating, but rather to combining, filtering, coalescing, and other phenomena. Viscosity of the oil is a factor in the precipitation of the water in oil, but this depends largely on the emulsion and on the temperature. Whatever dehydration methods are employed, the optimum temperature is determined from consideration of the gravity and viscosity, both as functions of temperature.

Temperatures from 160 to 250° F. are used in the majority of desalting installations, with as much as 350° F. used in isolated cases. Failure of some crudes to respond to heat and settling treatments resulted in the development of systems where the emulsion is passed through a contactor tower containing "hay" or excelsior, gravel, and more recently, fibre glass. In this type system the filter medium serves to coalesce the globules of water in the oil so that they will stratify due to gravity in a settling drum.

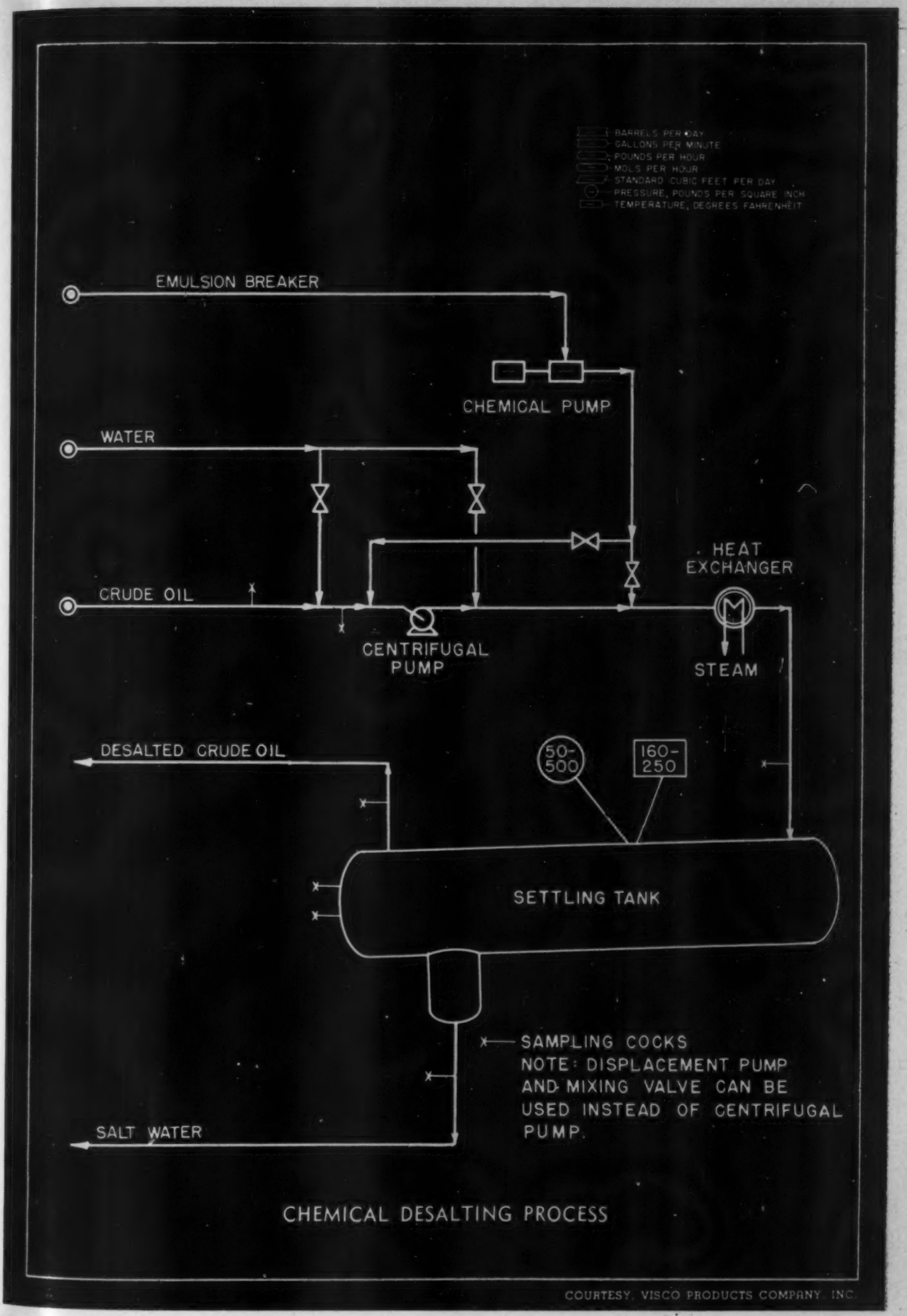
Chemical treating of emulsions for water and salt removal has progressed extensively and is now one of the simplest and most economical methods in use. Desalting is actually an extraction operation. The crude charge is emulsified with fresh water, and the resulting emulsion contains a less concentrated brine contamination which is subsequently removed by demulsification through chemical treatment and settling.

Several factors require careful attention for best results. A sample of the crude must be checked to determine the chemical that performs best in breaking the emulsion and gives the lowest cost per barrel of treated oil recovered. The chemical must be thoroughly mixed with the emulsion. The temperature selected must provide efficient coalescence and stratification of the water.

A typical refinery installation for chemical desalting of crude oil is shown on the flow diagram. Preheated fresh water varying from 3 to 10 percent by volume is added to the crude oil containing brine and is emulsified. Emulsification may be secured by flow through a charge pump or a mixing device. Emulsion breaking chemical is injected by a metering pump and the crude oil mixture flows to a heat exchanger where the temperature is increased to 180° F. or higher. The emulsion is broken at the heat exchanger and flows on to a settling vessel for separation. This vessel often is designed to allow settling time of one hour, and may be equipped with heating elements to maintain elevated temperatures and with baffles designed to reduce flow turbulence and channeling. The bottom of the settling drum may have a slope of about 2 percent toward a water trap to insure drainage.

Desalted crude oil leaves the top of the drum and salts dissolved in the fresh water are withdrawn from the bottom. Average treating cost will range from 0.5 mils to 2 mils per barrel of oil desalted. Initial investment is low.

The treating system and equipment in the accompanying flow diagram, while not licensed or patented, is one of the methods recommended for use in crude oil desalting operations in the refinery. Chemicals for crude oil desalting are manufactured and supplied by Visco Products Company, Inc., and by Tretolite Company. This discussion was prepared by Visco.



CHEMICAL DESALTING PROCESS

COURTESY, VISCO PRODUCTS COMPANY, INC.

Electrical Desalting Process

PETROLEUM RECTIFYING COMPANY

THE Petreco Electrical Desalting Process was developed by Petroleum Rectifying Company to provide an efficient and satisfactory method for removing salt and other solid materials from crude charging stock.

Salts in crude oils are usually carried by residual brine droplets which have been so highly stabilized by the emulsifying agents in the oil that they cannot be removed by the customary dehydration methods. Because of this, any refinery charging stock that contains any water is certain to contain some salt, the amount depending upon the water content (B.S.& W.) and the salinity of the brine.

The amount of salt entering the refinery equipment via the charging crude often reaches considerable proportions. For example: 10,000 barrels of a crude containing only 0.15 percent water will often carry close to half a ton of salts. The potential injury to the refining equipment from the continuous introduction of salts in such quantities can be visualized easily.

Salts other than chlorides are also present in oil field brines. In the past, refiners have been most concerned with the chlorides because of their ready conversion by hydrolysis to the very corrosive hydrochloric acid. However, salts such as sulphates and bicarbonates are often present in considerable proportions, and will most certainly contribute to the plugging of exchanger and furnace tubes. In evaluating a crude oil, a complete analysis is very desirable since a chloride analysis alone often will be inadequate.

Other impurities such as suspended solids of various origins are frequently found in crude oils. They may be sand and silt from the formation itself, drilling mud, rust, scale, etc. In PETRECO desalter installations, such solids are removed along with the salts, thereby preventing potential stoppages and erosion in the refinery equipment.

The process is carried on by adding fresh water to the salt-bearing raw crude. The added water is dispersed into the raw crude by the action of an air operated emulsifier. The emulsion so formed is subjected to the action of a high-intensity electric field, wherein the added water particles and brine particles are coalesced into masses containing the detrimental material. The process is continuous and can be tied into the refinery stream so that the oil is purified as it goes to the stills.

The volume of fresh water required ranges from 2 to 15 percent of the volume of oil being desalted —such volume being determined by the nature of the crude oil and the concentration of salts present therein. The pressure drop across the emulsifier varies according to the type of oil being processed—ranging from 3 to 75 pounds per square inch. The operating temperature for the process varies from 140° to 240° F. Since the process operates through high-tension electrostatic forces, the electric power consumption is low, the average load being 4 kw. per unit. After subjection to the action of the electric field, the oil and salt-bearing water separate and are drawn from the desalter in two continuous streams.

The standard Petreco desalter is a cylindrical pressure tank 10 feet in diameter by 12 feet on the straight side. The tanks are set up vertically, and including piping, are about 17 to 23 feet overall in height. Specially designed high-capacity units with throughput capacity up to 60,000 barrels per day are in operation. A number of the standard units can be installed as a battery; the limit to the capacity of the standard unit installation is dependent upon the complexity of piping that is practicable. Multiple units are manifolded together for automatic control of the water level. Electric power control is maintained by individual switch gear operating at either 220 or 440 volts.

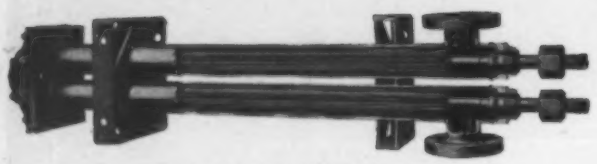
The Petreco process is in use throughout the United States, and in Canada and South America. Some of the oils desalted contain as high as 1100 pounds of salt per thousand barrels of oil; at the same time, many refiners desalt oil bearing as low as 10 to 15 pounds per thousand barrels. The Petreco process is highly efficient, frequently reducing the salts to values so low that the usual analytical methods are unable to detect them; there also is an accompanying reduction in all corrosive substances except sulfur. Effectively removed also is suspended insoluble matter such as sand, silt, volcanic ash, etc. Besides its normal function of removing salts and foreign solids, Petreco desalters provide a dependable means of removing slugs of water which, inadvertently, may find their way into the crude charge, and result in damage to the stills or interruption of operation.

The Petreco Desalting process is engineered, serviced and licensed by Petroleum Rectifying Company of California, Los Angeles.

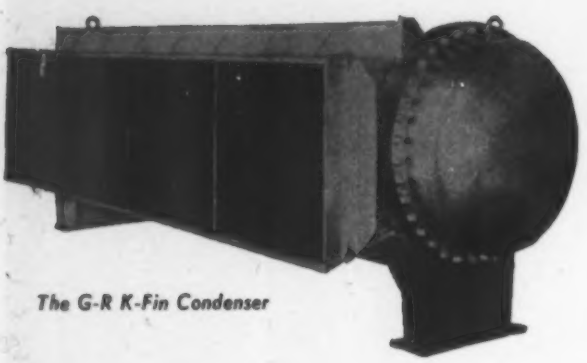
ELECTRICAL DESALTING PROCESS
COURTESY PETROLEUM RECTIFYING COMPANY
DESALTER
DESALTER
DESALTER
DESALTER
WATER BLEED SIGHT GLASS
EFFLUENT WATER TO SEWER
EMULSIFYER
WATER 60 200
CRUDE 140 240
WATER PUMP
LLC
CRUDE LEVEL
SURGE DRUM
STILL CHARGE PUMP
COLD UNTREATED CRUDE PUMP
PRIMARY CRUDE EXCHANGERS
SECONDARY CRUDE EXCHANGERS
TO FURNACE OR FRACTIONATOR
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

GRISCOM-RUSSELL

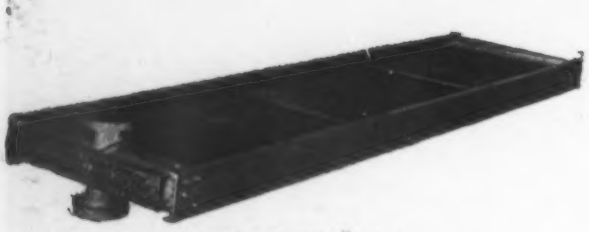
Pioneers in Heat Transfer Apparatus



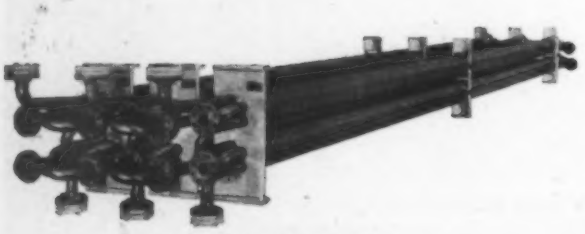
The G-R Twin G-Fin Section



The G-R K-Fin Condenser



The G-R Bentube Section



The G-R Tubeflo Section

G-R Tubular Heat Exchanger

The Griscom-Russell Co. has continuously pioneered in the design and application of heat transfer apparatus for 79 years. The unequalled experience of the Company in this field assures authoritative recommendations and an exactly suitable design for every requirement of heaters, coolers, condensers and heat exchangers to handle liquids, gases, vapors and fluid solids.

The Company has always been in the forefront of the origination and development of special types of surfaces to provide the most practical and effective heat transfer for special conditions and services in which conventional types of shell-and-bare-tube apparatus are inefficient or troublesome. The principal types of distinctive G-R heat transfer surfaces are briefly outlined below:

G-FIN APPARATUS

Heat transfer elements consist of pipes or tubes with longitudinal fins which provide six to eight times as much external surface as on bare tubes of the same diameter. The fins therefore compensate for low heat conductivity of a fluid to be heated or cooled by another fluid of higher conductivity. More than 50,000 installed G-Fin units have proven the effectiveness and serviceability of this patented design.

K-FIN APPARATUS

These helical finned G-R heat transfer elements are particularly adapted for services where a fluid of low heat conductivity flows transversely across the tubes. This design provides the only practical units for condensing large quantities of vapor at high vacuum when mixed with steam which is not to be condensed. The K-Fin surface is also particularly adapted to cooling or condensing units where air is to be the cooling medium.

BENTUBE SECTIONS

Heat transfer elements set with a slight initial curvature into cast headers held rigidly in position. Changes in temperature therefore produce corresponding changes in flexure of the elements, causing scale formations due to hard or salt water to crack up and flake off. G-R Bentube Sections are widely used for condensing hydrocarbon vapors and cooling petroleum derivatives; and similar scale-shedding elements are also used in the well-known G-R Bentube Evaporators for distilling boiler feed make-up.

TUBEFLO SECTIONS

This design consists of four parallel tubes expanded by a patented process into closely spaced diamond shaped fins. Each fluid flows through two of the tubes and the fins transfer the heat from the hot tubes to the cold tubes. This arrangement provides the only design of closed type heat exchanger that will maintain rated capacity when handling dirty or corrosive fluids and also makes impossible leakage of one fluid to the other fluid.

TUBULAR HEAT EXCHANGERS

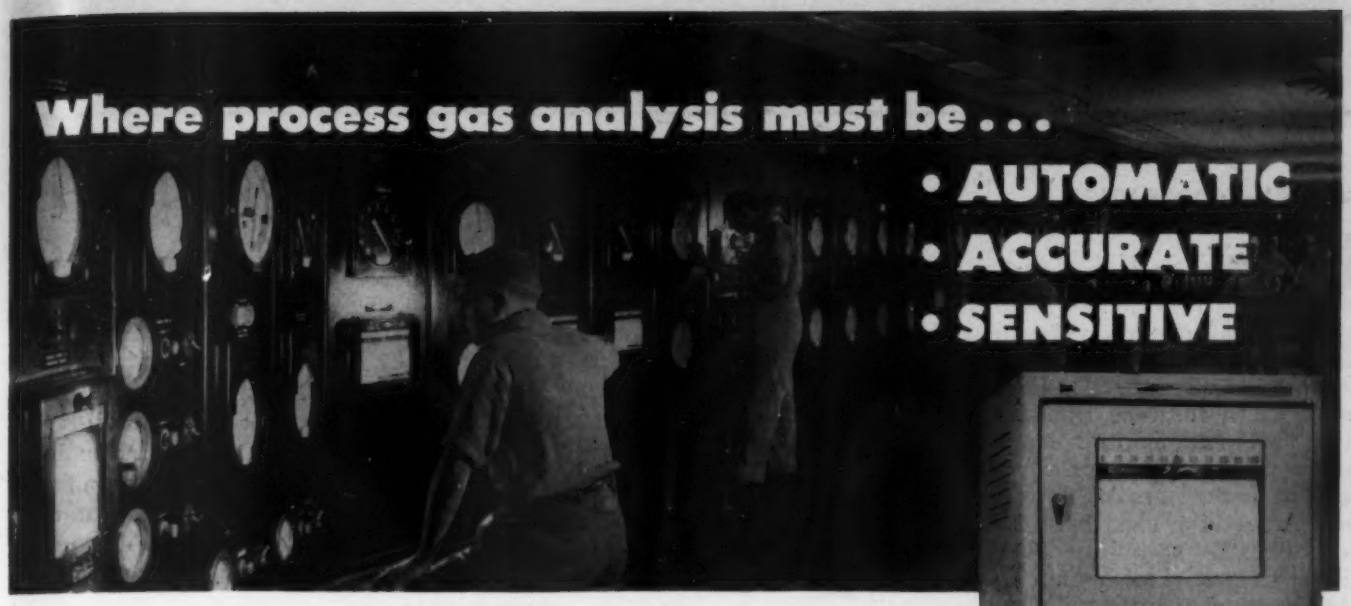
G-R designs of heat transfer apparatus also include a wide variety of types of "bare-tube" units for all requirements. The extensive range of G-R equipment permits impartial recommendations, and the long experience of the G-R engineers assures authoritative advice on your heat transfer problems.

Bulletin describing suitable equipment for your particular requirements will gladly be sent on request.

THE GRISCOM-RUSSELL COMPANY

285 Madison Ave., New York 17, N. Y.

Sales Offices and Representatives in Principal Cities



THE *Baird Associates* Infrared Absorption Type AUTOMATIC RECORDING GAS ANALYZER

. . . offers these important advantages for control of catalytic cracking, synthesis of fuels, and many other refining processes—increasing both operating efficiency and productivity:

- **Fully automatic process control**

  The gas analyzer unit automatically maintains process control for a critical concentration of the specified gaseous component by its ability to govern a standard pneumatic system operating at the control point.

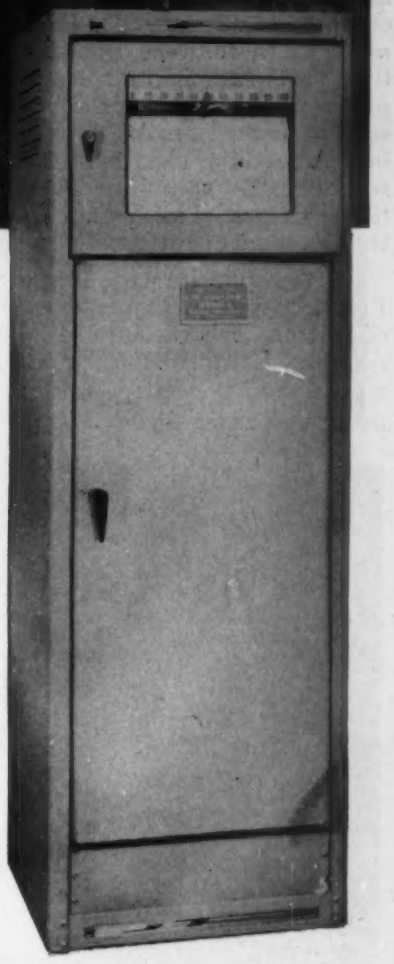
- **High sensitivity with high discrimination**

  Ultimate sensitivity for most gases is approximately 10-20 parts per million. For gas mixtures where infrared absorption bands overlap, an additional "interference cell" minimizes the interfering components to give high discriminating power.

- **Operated and maintained by plant technicians**

  Plant technicians assume the entire burden of chemical analysis of process gases, relieving highly skilled personnel of an expensive and time-consuming task.

- **Thermostatic controls** prevent temperature errors.
- **Continuous automatic standardization** of "zero drift" takes place hourly.
- **Sealed-in reference gas** unit insures reliability.



The Baird Associates infrared gas analyzer can be furnished in styles to suit various requirements, including remote indicating and recording, explosion-proof housings, and the cabinet style shown above.

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For complete details on the Baird Associates infrared gas analyzer, request Bulletin XXXI. For information on its specific application to your problems, we will gladly study your specifications and report.

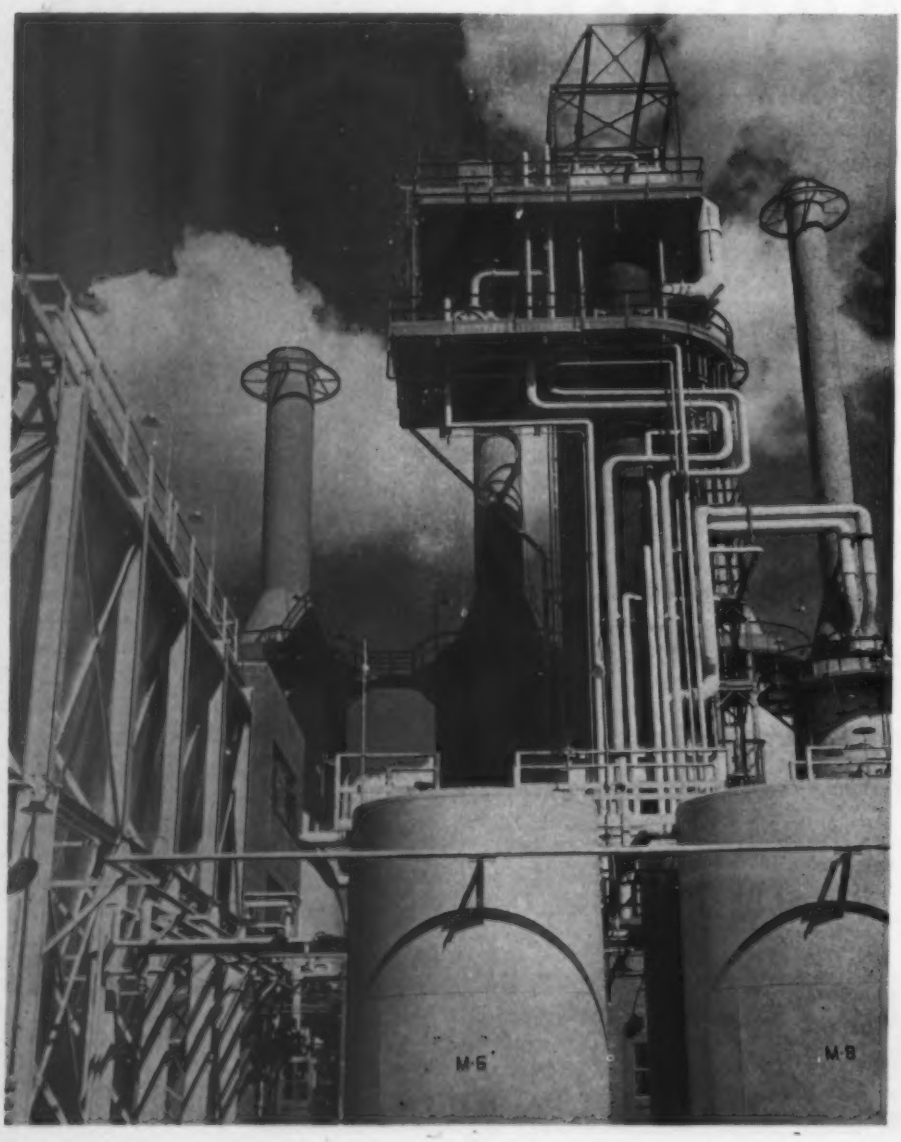
Baird Associates, Inc.
INDUSTRIAL PHYSICISTS
33 UNIVERSITY RD. CAMBRIDGE 38, MASS.

Filtrol Fractionation Process

FILTROL CORPORATION

FILTROL Fractionation is the name of the process in which Filtrol, the highly activated adsorbent made by Filtrol Corporation, is injected into a hydrocarbon oil stream as it flows to fractional distillation equipment. The purpose of this relatively new technique is to combine fractional distillation and continuous filtration into a single and more economical process. The process is applicable to the refining of solvent-treated stocks, long residuals, low-asphaltic or treated lube crudes, and dewaxed cylinder stock solutions. By combining distillation and decolorization, substantial savings in equipment, upkeep, maintenance and operating costs are effected.

On the opposite page is a flow diagram of a Filtrol Fractionation plant presented through the courtesy of Filtrol Corporation, 634 South Spring Street, Los Angeles, California. As indicated in the flow sheet, a portion of the charge to the pipe heater is continuously by-passed through a small mix tank. This tank is equipped with a mechanical agitator which mixes the Filtrol with the oil, thus forming a readily pumped oil-adsorbent slurry. The Filtrol is automatically fed to this tank by a gravimetric feeder, insuring accurate control of the adsorbent feed at all times. The feeder is supplied with Filtrol by a scroll-type conveyor from the overhead storage bin. The by-passed oil containing the Filtrol rejoins the main flow of the charge and continues through the heat exchangers before entering the pipe heater. The temperature of the heater outlet is instrument controlled. The hot oil-adsorbent slurry then enters the vaporizer section of the vacuum fractionating column where the unvaporized portion which contains all the Filtrol falls into the bottom soaking section. The lighter cuts thus receive an adsorbent treatment in passing through the exchangers and furnace, after which the adsorbent charge is concentrated in the bottoms, which require heavier treatment. The vaporized material is fractionated into the desired cuts, each of which passes through a steam stripper which eliminates undesirable odor and insures color-stable products.



The bottoms and Filtrol in the soaking section of the tower are agitated and stripped by the injection of steam into the bottom of the vessel. The soaking time or contact time is regulated by controlling the liquid-level in this section. The slurry then passes through the bottoms exchangers and into an insulated surge tank from which a continuous-vacuum rotary filter is fed. The spent adsorbent is filtered from the oil which passes through a cooler to finished oil storage. The spent cake is continuously removed by a screw conveyor.

In certain instances, the Filtrol Fractionation Process has definite benefits over contact filtration finishing. The refiner's position, process- and market-wise, must be carefully analyzed to determine the extent of these advantages for each specific case. Important advantages realized by the use of this refining technique include reduced operating cost, lowered construction cost, and maximum adsorbent economy.

FILTROL FRACTIONATION PROCESS
COURTESY, FILTROL CORPORATION
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
FLUE GAS HOLDER
FLUE GAS FROM GENERATOR
EDUCTOR
STEAM
TO SLOP
PRECOAT OIL
SPENT FILTROL BIN
PRECOAT TANK
SLOP TANK
FILTER
FILTER SURGE TANK
O H GAS OIL
BRIGHT STOCK
VACUUM TOWER
50 MM ABS PRESS
HEATER
AIR ASPIRATOR
MIX TANK
FEED BIN
GRAVIMETRIC FEEDER
CHARGE OIL
NO 1 SIDE STREAMS
NO 2 NO 1

View of Oceco 3" Screwed Connection Vent Valve with Flame Snuffer and Chain.

OCECO
V-113 VENT VALVES
for maximum efficiency, economy and dependability

View of Oceco 6" Flanged Connection Vent Valve showing mounting on Oceco Flame Arrestor.

● Oceco Vent Valves provide positive and dependable control of tank breathing. They prevent the free discharge of vaporized liquids resulting from fluctuations in temperature,—and restrict tank breathing, resulting from pumping into a tank. They avoid any serious lowering of gravity ratings;—retard the formation of gum;—and avoid "cave-ins" and "blow-outs" resulting from excessive vacuum and pressure.

These valves combine the advantage of our twenty five years of experience in meeting the industry's most exacting requirements;—and the results of exhaustive tests conducted in our laboratories. The Venturi effect of the body gives the valves even greater pressure capacity than an open nipple, while the spacing and design of the openings provides large vacuum relief. The drip edges on the valves carry away any condensate, thus protecting the valves against corrosion and freezing, and the stem guided construction prevents sticking and "cocking."

The housings are semi-steel castings that can withstand direct exposure to flames for long periods of time, and consequently provide a solid support for the valves, valve guides and seats, keeping them in proper alignment and assuring tight closure at all times.

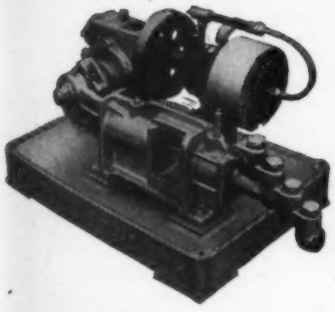
Furnished in 2" and 3" sizes for screw mounting, and 4", 6", 8", 10" and 12" sizes for flange mounting, either as individual units or complete with an Oceco Flame Arrestor. Prices and a fully descriptive bulletin sent promptly on request.

OCECO
Division of THE JOHNSTON & JENNINGS CO.
CLEVELAND 14, OHIO

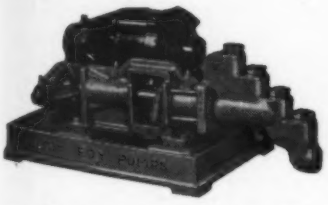
NEW YORK • PITTSBURGH • DETROIT • CHICAGO • ST. LOUIS • TULSA • BEAUMONT
MOBILE • CASPER • SAN FRANCISCO • LOS ANGELES • TORONTO AND MONTREAL

MILTON Roy PUMPS

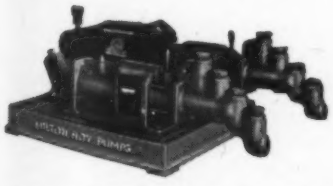
Automatic Chemical Feed Systems for Metered Chemical Feeds



Simplex Pump with Rotary Air (or Gas) Motor Drive.

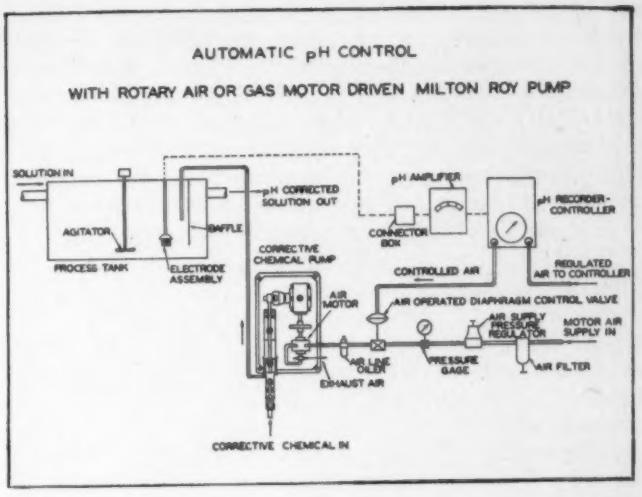


Standard Simplex Pump, 1 pt. to 300 gph. Pressures up to 6000 psi.



Standard Duplex Pump, 1 qt. to 600 gph. Pressures up to 6000 psi.

Catalog 146 describes above and other Milton Roy Standard and Special Pumps also Automatic Control Systems



In refinery operations hundreds of Milton Roy Pumps are in successful use handling hot or cold distillates, solvents, furfural, sweet or sour naphtha fractions, liquefied petroleum gases, tars and asphalts, acids and acid sludges. They are used in solvent extraction, blending processes and controlled catalyst make-up. For pilot plant applications they combine in a single unit a metering and liquid transfer medium against any desired pressure. For use where electric current is not available a Milton Roy Pump powered by a rotary air (or gas) motor is offered.

These Milton Roy Pumps are now available with completely automatic volume control, for applications such as adding inhibitors, neutralization of sulfides or others where very precise automatic metering of the volume of liquid pumped may be advantageous. This precise control of volume, down to 1 pint per hour, is accomplished by means of electronic circuits which regulate the speed of the pump (strokes per minute) in direct ratio to any variable that can be metered . . . such as rate of flow, temperature, pressure, liquid level, gas analysis, chemical concentrations, etc.

The electronic control is simple, compact and rugged . . . employing standard long-lived units. Pumps are positive displacement units of standard Milton Roy Step-Valve design, in practically any metal and some plastics construction, to meet most corrosion problems, to pump up to 23 gals. per min. per pump valve, ar.[illegible] against pressures up to 20 thousand psi. Automatic systems are supplied complete with pumps, meters, control units, etc., as may be required.

Consult with Milton Roy engineers. Ask for new Catalog 146, also Technical Paper No. 54 "Electronics in Automatic Chemical Feed Systems." Other available current literature includes Bulletin 451 "Milton Roy Chemical Feed Systems for Boiler Water Treatment," Technical Paper No. 51, "What the Power Engineer Needs to Know About Chemical Feed Systems," Technical Paper No. 52 "Pulp Digestion," Technical Paper No. 53, "Soap Manufacture," Technical Paper No. 55 "Automatic pH Control in Water and Industrial Waste Treatment," Technical Paper No. 57 "Controlled Volume Chemical Pumps, Design and Operation."

MILTON ROY COMPANY

1378 E. Mermaid Lane, Philadelphia 18, Pa.

See your classified 'phone directory for local representative

Percolation Filtration Process

ATTAPULGUS CLAY COMPANY

THE percolation filtration process for finish-refining of petroleum fractions is used to produce lubricants, distillates, waxes and specialty-oil products to the high-quality specifications. Percolation-finished oils and waxes are characterized by excellent color, odor and taste, and by their stability and demulsifying properties. In recent years, improvements in equipment and more rigid application of technical control to filter-plant operation have increased processing efficiency and decreased operating costs.

Percolation filtration consists essentially of the filtering of a partially refined petroleum fraction through a columnar bed of granular adsorbent material. The adsorbents commonly used are fuller's earth and activated bauxite. Each of these materials possesses certain inherent refining advantages, depending upon characteristics of the oils being finished, and selection of the proper adsorbent is therefore the first requisite to percolation-plant design.

The mechanics of the percolation process are such that the oil and adsorbent are brought together under conditions which approach true counter-current operation. The finished products receive thereby the most thorough degree of purification.

Percolation adsorbents are revivified for reuse by burning away of the impurities adsorbed from the oils.

Lubricating-oil stocks may be refined by this process by percolation in dilution with 50 percent or more of naphtha or by the straight percolation of the oils without such dilution. Solvent-refined oils wherein the color reduction required in percolation finishing is relatively small and which are filtered primarily for stability, demulsibility, etc., usually may be filtered without dilution to high yields of finished product. A flow plan and general considerations of a plant of this latter type, suitable for the finishing of solvent-refined oils or waxes, is shown on the opposite page through courtesy of Attapulgus Clay Company, 260 South Broad Street, Philadelphia 1, Pa.

The adsorbent used in percolation finishing of mineral oils and waxes is usually a 20/60 or 30/60 mesh. The new adsorbent is heated in the adsorbent burner to 600-700° F. prior to first use. The dried or "tempered" adsorbent then is charged to the filter which may vary in capacity from 10 to 50 tons. The filter then is filled with oil and after a short soaking period the percolation proper is started. The first oil off is usually lighter colored, the stream gradually darkening until the refining power of the adsorbent may be considered to be exhausted. The filter then is drained, washed with naphtha, steamed to recover the naphtha and to condition the adsorbent for revivification. At the completion of this part of the cycle, the adsorbent is drawn from the filter and revivified in the burner at a maximum temperature of 1100° F.

The decolorizing and refining efficiency of granular fuller's earth is gradually decreased by the revivification treatment and the usual practice is to replace the plant charge at intervals. The proper point of clay discard varies with the type of oil stock being processed and is determined by cost calculations. The number of revivifications economically obtained may vary from 8 to 18 with average operating conditions and equipment.

The revivification characteristics of activated bauxite differ in several respects from fuller's earth, the most significant difference being its greater resistance to loss in efficiency caused by the revivification operation. This smaller efficiency decline obviates the necessity for periodic discard and it becomes possible to operate a filter plant under stabilized adsorbent efficiency conditions, the bolstering effect of the makeup material necessary to compensate for unavoidable handling losses just balancing the small decline in efficiency caused by revivification. The equilibrium level thus established will vary with the type of oil being treated and type of equipment used. It usually will fall within the range of 60-80 percent of the efficiency of new bauxite.

The advantages of fuller's earth are particularly evidenced in the manufacture of white medicinal oils and in the treatment of lubricating stocks where the desired color reduction is particularly high. Activated bauxite, on the other hand, displays advantageous qualities for the refining of petrolatums and waxes to light colors and to odor and taste-free specifications. The greater resistance of bauxite to efficiency degradation frequently will recommend its adoption from economy considerations. The refining ability of bauxite is useful especially in the treatment of Pennsylvania residual stock and in the finishing of special lubricating oils to high-quality demulsibility requirements.

Owing to the variable properties of oils from different crude sources and varying pre-treatments, the selection of the proper adsorbent for a given decloorizing problem should be the aim of a detailed investigation and study of the oils to be treated.

Regardless of the adsorbent employed, the percolation-filtration process possesses certain inherent merits which should be borne in mind in considering the installation of decolorizing facilities.

It is essentially a low-cost process, the greater part of the manufacturing cost comprising expense other than adsorbent cost. For this reason improvements in operating procedures are reflected in reduced decolorization costs to a much greater degree than other processes in which the adsorbent cost constitutes the major item of expense.

It is essentially a counter-current process in contrast to other decolorization methods which frequently represent the least desirable application conceivable from the standpoint of counter-current operation.

It is flexible, permitting simultaneous processing of several products in the filter house as well as production of several "specialty cuts" from the same filter.

It is an adsorption process operating at low temperatures and, as a result, yields finished oil of generaly higher quality.

Ultimate yields, that is, yield of decolorized oil in percent of stock charged, are generally high.

A recent development in percolation filter plant design has been the employment of an open plan of construction in place of the conventional walled-in filter house. This has materially decreased installation costs and has also eliminated a number of safety hazards long inherent to the older design.

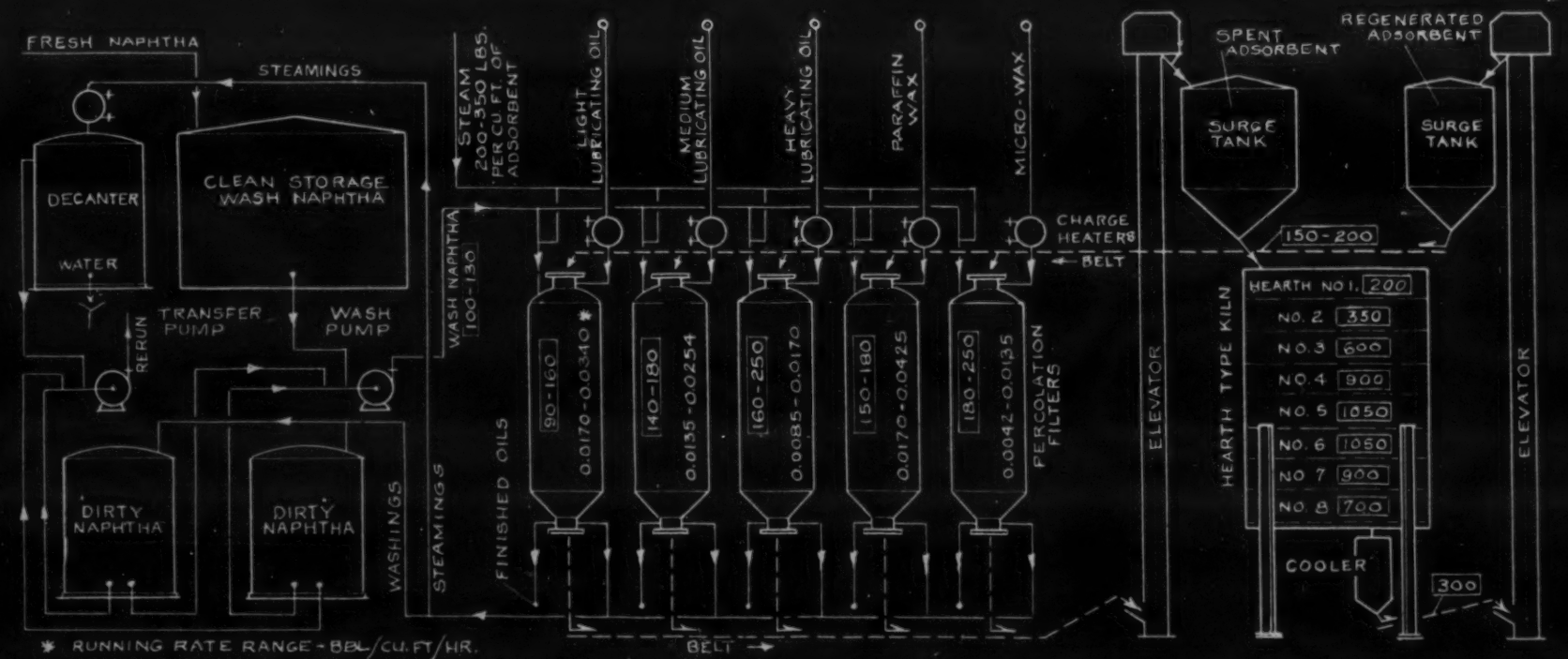
TYPICAL YIELDS

LUBRICATING OILS

Crude Source	Processing	Vis. SU 210° F.	Color ASTM	Adsorbent	FILTRATE Color	FILTRATE Bbl. Ton
Louisiana	Phenol	45.7	5¼	Bauxite Fullers Earth	1½ 1½	43.0 67.5
Penna.	200 Neutral	47.0	2+	Bauxite Fullers Earth	1½ 1½	48.0 76.6
California	S.A.E. 20 Phenol	51.5	4-Dil.	Bauxite Fullers Earth	4 4	126.0 156.5
Oklahoma	Furfural	62.6	4¼	Bauxite Fullers Earth	2 2	113.0 52.0
Gulf Coast	Duo-Sol	119.3	7+	Bauxite Fullers Earth	6 6	60.0 62.8
California	S.A.E. 60 Phenol	120.2	4½	Bauxite Fullers Earth	4 4	365.0 550.0

WAXES

Crude Source	Processing	M.P. °F.	Color	Adsorbent	FILTRATE Color	FILTRATE Bbl. Ton
California	Propane-M.E.K.	143.2	½ N.P.A.	Bauxite Fullers Earth	30 Say.	250.5 331.0
Oklahoma	M. E. K.	122.4	1+ N.P.A.	Bauxite Fullers Earth	25 Say.	17.6 11.3
Illinois	M. E. K.	131.0	2¾ N.P.A.	Bauxite Fullers Earth	20 Say.	37.0 32.9
Mid-Continent	M. E. K.	174.8	Brown	Bauxite Bauxite	½ N.P.A. 1¾ N.P.A.	7.0 15.9
Oklahoma	Unknown	185.0	Black	Bauxite Bauxite	18 Say. 1 N.P.A.	6.3 11.4
East Texas	M. E. K.	166.5	7 N.P.A.	Bauxite	1 N.P.A.	21.6
California	Propane-M.E.K.	158.6	2¾ N.P.A.	Bauxite	18 Say.	105.2



PERCOLATION FILTRATION PROCESS

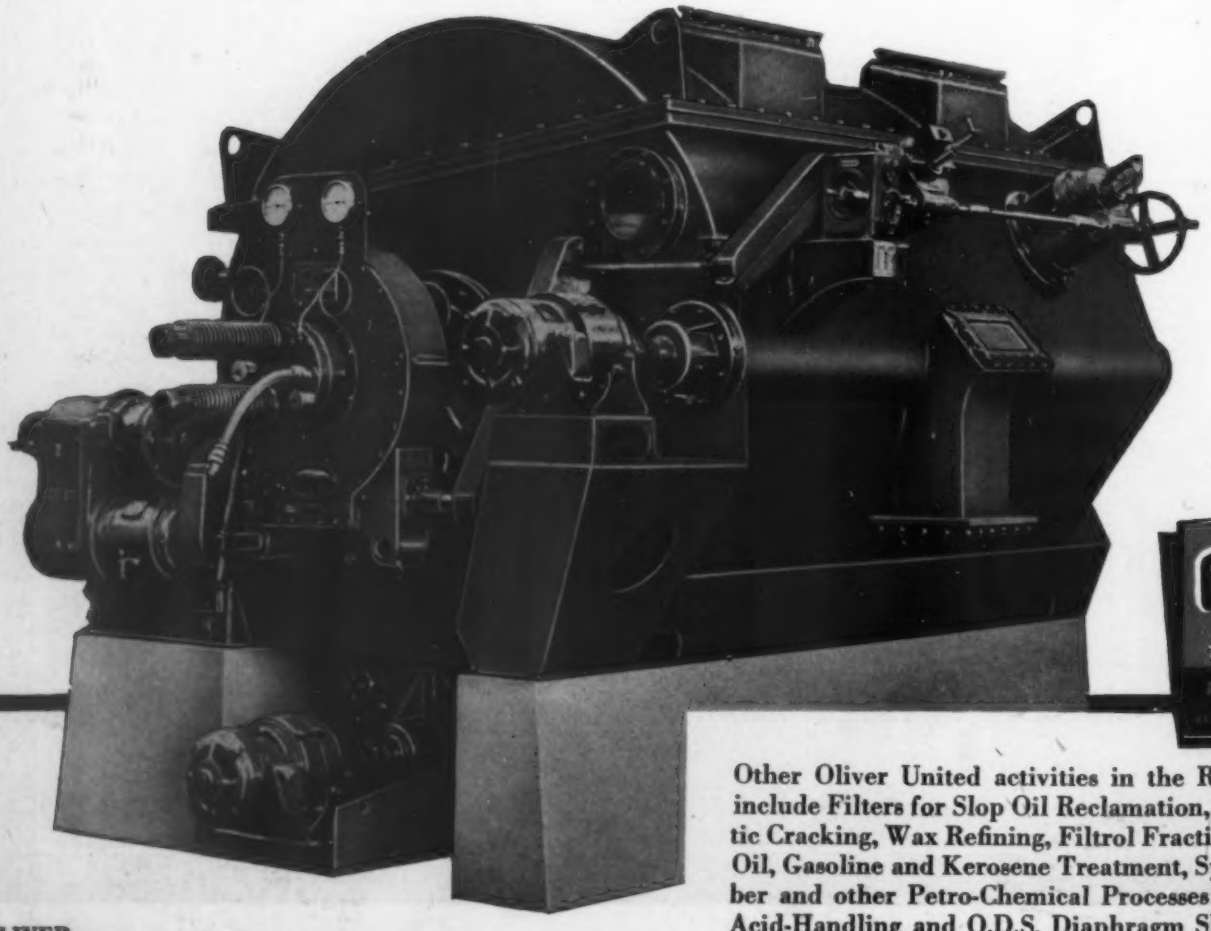
COURTESY, ATTAPULGUS CLAY COMPANY

CONTACT CLAY

Why Handle Contact Clays ***Manually*** ???

Why Continue to ***Lose*** *Too Much Oil* ???

Why Operate in ***Batches*** *with High Maintenance* ???



OLIVER
CONTINUOUS VACUUM PRECOAT FILTER

Other Oliver United activities in the Refinery Field include Filters for Slop Oil Reclamation, Fluid Catalytic Cracking, Wax Refining, Filtrol Fractionation, Fuel Oil, Gasoline and Kerosene Treatment, Synthetic Rubber and other Petro-Chemical Processes; also Olivite Acid-Handling and O.D.S. Diaphragm Slurry Pumps. All are described in Bulletin 405-R.

REMOVAL !

. . . The OLIVER PRECOAT FILTER says:

"You Shouldn't; You Don't Have To."

It's a significant record, the one showing that *all new* domestic Lube Oil Contact Plants are being equipped with Oliver Precoat Filters for clay removal! It's also a matter of record that wherever existing contact plants have switched from batch clay removal in pressure filters to continuous removal with the Oliver Precoat, the Precoats have paid for themselves in a very short time . . . most of them in less than a year. And the trend is accelerating: *continuous* Oliver Precoats superseding *batch* Sweetlands.

Savings with the Oliver Precoat Filter can be traced directly to these main advantages:

1 . . . oil losses in spent clay are cut from 40 to 50% content to around 20 to 25%; or to put it another way, savings range from 50 to 60%.

2 . . . shifting from manual batch to continuous automatic handling saves from 50 to 75% in labor cost.

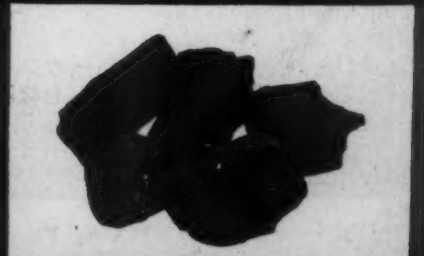
3 . . . and there's considerable saving in maintenance cost largely because periodic replacement of individual leaves is eliminated; and extra leaves need not be kept in stock.

As a famous public official once said: "Let's look at the record." Every new domestic Lube Oil Contact Plant since 1937 has been equipped with Oliver Precoat Filters. That means something. If you are operating clay removal filters of the pre-precoat era, ask to have one of our representatives call to give you a report on the savings to be had by switching over to the Oliver Precoat Filter.

OLIVER UNITED FILTERS INC.

New York 18, N. Y.
33 West 42nd Street
San Francisco 11
California
Chicago 1, Ill.
231 S. LaSalle Street
Western Sales Division
Oakland 1, Calif.
2900 Glascock Street

Typical Precoat Filter cake discharge, readily handled on any kind of mechanical equipment.



Pressure Filter Cake Discharge, difficult to handle because of much higher oil content.

Continuous Contact Filtration Process

FILTROL CORPORATION

CONTACT filtration is a process wherein petroleum lubricating oils are rapidly and efficiently refined by the simple procedure of mixing oil directly with Filtrol, the activated adsorbent made by Filtrol Corporation, heating for a short period and then separating the decolorized oil from the spent cake.

The superiority of this process over other methods of refining has been increased by recent improvements in equipment design and in bleaching adsorbents, which in turn have simplified the operation, improved product quality and lowered costs. Some of the major advantages which are now recognized are briefly summarized below:

Continuity of Operation. The contact process as now developed permits continuous operation from start to finish. This advance has been accomplished largely as a result of recent improvements in equipment manufacture.

Large Throughput with Minimum Space and Equipment Requirements. Inasmuch as this process provides ready adjustment of contacting temperature it is obviously possible to handle stocks of a wide viscosity range with but little effect on plant throughput. The cost of modern contact-plant installations is relatively low. Such a plant will occupy a space not greater than 80 feet wide by 120 feet long.

Ease of Switching from One Type Stock to Another. Due to the flexibility of the contact equipment and its small volume, changes from one type oil to another can be effected with minimum time loss.

Low Labor Cost. A modern 2000-barrel-a-day contact plant can be operated with two men per shift. Equipment design has undergone improvement to the extent that substantially each unit of the plant can be controlled automatically.

Rapid and Accurate Cost Analysis. Simplicity of the contact process makes it possible to obtain accurate daily cost data. Thus the management is constantly provided with true costs of the finished oils, either individually or collectively.

High Quality Products. Uniform and predictable colors, superior casts, lower carbon residues, improved emulsion and oxidation tests, are among the outstanding merits of this process. The use of fresh and uniformly standardized adsorbent in the contact process eliminates the uncertainties.

Adaptability to Other Refinery Processes. A contact plant can be used to clarify, decolorize, and/or neutralize lubricating oils whether they have been acid-treated, solvent-extracted or simply distilled. In addition to finishing the normal production of such processes, the contact plant often is used to correct previous faulty refining. Sour oils are neutralized and decolorized in one operation, eliminating caustic neutralization and the accompanying disadvantages of emulsion formation and poor casts.

Low Operating Costs. In a modern contact-filtration unit where adsorbent handling is accomplished by an automatic mechanical system, the operation cost averages approximately $0.08 per barrel for a 2000-barrel-per-day plant.

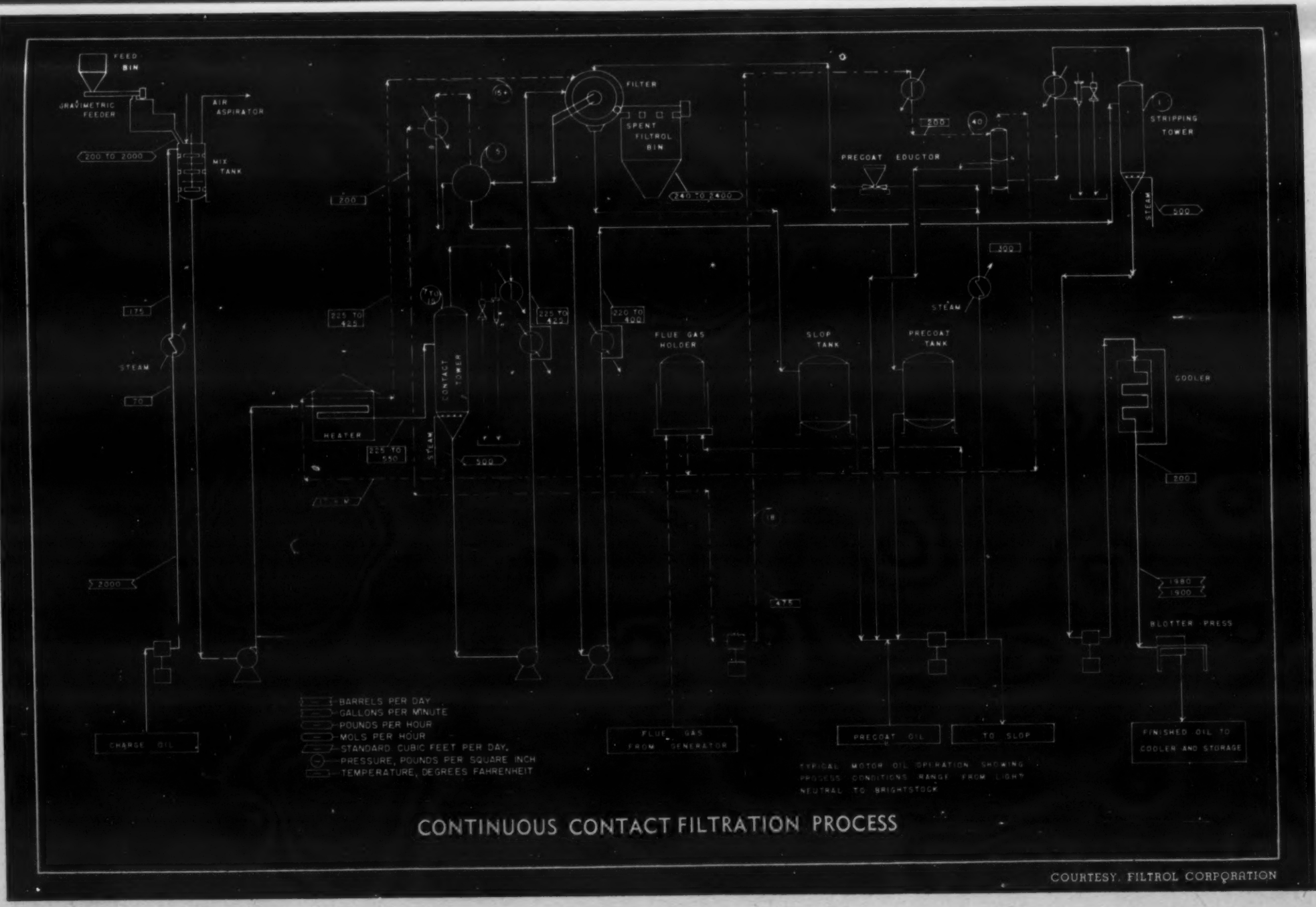
Maximum Adsorbent Economy. The flexibility of modern design in a contact filtration unit combined with improved adsorbents enables the process to be operated under optimun conditions. This positive control of mixing, heating, filtering and stripping results in maximum adsorbent economy. Generally speaking, activated adsorbent efficiencies increase with temperatures, the optimum point being at or near the flash point of the contact oil.

On the opposite page is a flow diagram of a continuous contact plant presented through the courtesy of Filtrol Corporation, 634 South Spring Street, Los Angeles, California. After Filtrol has been added by means of automatic proportioning equipment to the oil change, the oil-Filtrol mixture enters a pipe-still heater where the maximum contact temperature is obtained in a once-through operation. From the heater, the slurry of oil and Filtrol enters a baffled vacuum stripping tower where a level is maintained regulating the time of contact. Steam sometimes is passed up the tower to facilitate the stripping action. The slurry is drawn continuously from the bottom of the tower, passed through the heat exchanger utilized to heat the fresh-oil charge, and thence to a filter surge tank which acts as a reservoir to feed the continuous precoat rotary filter. The filtered oil from this filter sometimes is charged to a high-vacuum stripper to furnish additional control of flash, odor, oxidation, and emulsion characteristics. The filtered oil finally is pumped through a blotter press which furnishes a polishing step before the oil is transferred to storage. In a few cases where the conditions warrant it, the spent filter cake is extracted with naphtha to recover the small additional quantity of oil soakage.

Filtrol addition, temperatures, rates, etc., are all subject to automatic control in modern installations, hence long uniform runs are currently made without difficulty.

...y of ...with ...rated ...mix- ...mum ...l ad- ...opti- ...con-

...con- ...sy of ...s An- ...neans ...ange, ...where ...once- ...of oil ...lower ...con- ...facili- ...ntinu- ...rough ...large, ...reser- ...The ...to a ...ol of ...The ...press ...trans- ...itions ...phtha ...kage. ...l sub- ...hence ...culty.



CONTINUOUS CONTACT FILTRATION PROCESS

COURTESY, FILTROL CORPORATION

USE THESE BETTER

Vogt Products

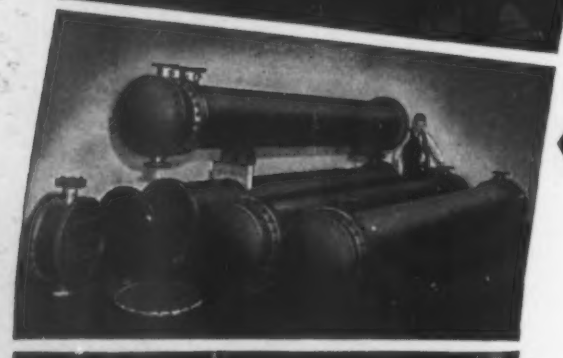
For Today's Tough Service Demands

EFFICIENT AND DEPENDABLE STEAM GENERATION

Vogt steam generating units are designed to give maximum rating in a minimum of space with high efficiency and low maintenance expense. Bent tube types and straight tube, forged steel sectional header types to burn solid, liquid or gaseous fuels, as desired, meet every power, heating or process requirement.

PROCESS EQUIPMENT FOR EVERY SERVICE

Stills, towers, oil chilling machines, filter presses, heat exchangers, etc. are constructed to all Codes. They meet all demands for operating security and trouble-free performance and help to lower costs in important process industries around the world.



SPECIAL MATERIALS FIGHT CORROSION AND PRODUCT CONTAMINATION

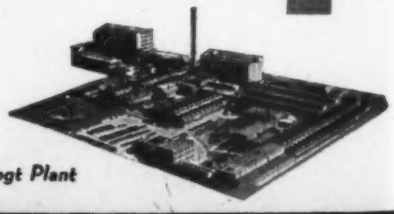
Our modern shops produce a wide variety of equipment made from special metals and alloys to combat corrosion, and product contamination or discoloration. Fabrication procedures employed insure that corrosion resistant properties of welds will match that of the materials from which units are constructed.

MORE TONNAGE AT LESS COST

Over 60 years of manufacturing experience, engineering and research stand behind Vogt refrigerating and ice making machinery. Absorption Systems, Compression Systems, and the Automatic Tube-Ice Machine in a wide range of capacities serve in leading petroleum refineries, chemical plants, ice and cold storage plants, dairies, packing plants, etc., at home and abroad.

DROP FORGED FOR EXTRA TOUGHNESS AND LONG-TIME SERVICE

Vogt valves, fittings and flanges, for top performance in oil, water, air, gas, and ammonia services, at high or low pressures and temperatures, are available drop forged entirely from carbon steel or stainless steel. Valves can be furnished in a combination of materials by using stainless steel for parts affected by service temperature or corrosion, and less expensive alloys or carbon steel for other parts.



Air View of Vogt Plant

HENRY VOGT MACHINE CO.

LOUISVILLE 10, KENTUCKY

BRANCH OFFICES: NEW YORK, PHILADELPHIA, CLEVELAND, CHICAGO, ST. LOUIS, DALLAS

YARNALL-WARING COMPANY

Manufacturers of YARWAY Steam Specialties

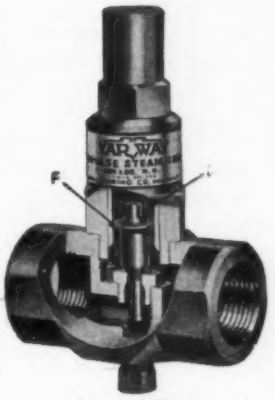
Trade-Mark

Main Office and Works: 140 MERMAID AVE., PHILADELPHIA 18, PA.

THE YARWAY IMPULSE STEAM TRAP

600,000 sold. Stocked by nearly 150 mill supply houses in key cities.

Simple construction. Made entirely of bar stock; body and bonnet of cold drawn steel, cadmium plated for protection against rusting; valve, seat and control cylinder of stainless steel, valve and seat heat-treated. Bronze capped traps for pressures up to 400 lbs., all-stainless-steel for 400 to 600 lbs.



Impulse Trap

Good for all pressures without change of valve or seat. Yarway Impulse Traps are factory set to operate at all pressures without adjustment or change of working parts. Six sizes, ½ to 2 in., serve nearly all trap needs.

Only one moving part, the valve (F) shown above. Valve and seat of stainless steel, heat-treated for long life.

Small size, hardly larger than a pipe union. ½-in. trap is 2⅝ in. long, 2-in. trap 4¾ in. long.

Light weight, ½-in. trap weighs only 1⅜ lbs., 2-in. trap weighs only 8⅝ lbs. No support other than the pipe line.

Low cost. Frequently cheaper to install new Yarway trap than to repair old ordinary trap.

Easily installed. Straight line connection simplifies installation. Removal of bonnet permits inspection of working parts without taking trap out of line.

High efficiency. Large air and condensate capacities. Fast heating and sustained efficiency.

Write for Catalog T-1739

OPERATION

Small portion of condensate is bypassed continuously around piston of valve F into controlled chamber K, discharging through orifice in center of valve into outlet side of trap. At low condensate temperatures, discharge through center orifice reduces chamber pressure and valve is lifted for full discharge through main seat. At near-steam temperature, condensate vaporizes at reduced pressure in control chamber. Expansion of this vapor raises chamber pressure and valve is closed, shutting off main discharge.

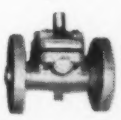
Under small condensate load, discharge is handled by control orifice and valve does not lift. At medium condensate loads, valve opens and closes at short intervals. Under heavy condensate load, valve opens wide and discharge is heavy and continuous.

CAPACITIES*, PRICES, WEIGHTS & DIMENSIONS

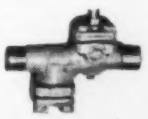
*Maximum continuous discharge of condensate in pounds per hour. Nos. 60 to 67 for 400 lbs. press., Nos. 120 to 127 for 600 lbs.

Pressure Pounds	½" No. 60 No. 120	¾" No. 61 No. 121	1" No. 63 No. 123	1¼" No. 64 No. 124	1½" No. 66 No. 126	2" No. 67 No. 127
10	475	825	1360	1910	2400	3425
20	650	1150	1725	2315	3050	4340
30	800	1375	1985	2650	3525	5050
40	900	1575	2220	2950	3950	5615
50	980	1725	2425	3200	4315	6100
60	1050	1850	2600	3425	4630	6525
70	1140	1975	2775	3660	4940	6930
80	1200	2080	2925	3875	5200	7260
90	1240	2175	3075	4075	5440	7600
100	1290	2260	3215	4280	5675	7875
125	1400	2460	3550	4760	6165	8530
150	1490	2640	3870	5175	6630	9075
200 to 400	1630	2900	4380	5960	7410	9950
Trap List Prices	$15.00	$22.00	$31.00	$48.00	$68.00	$90.00
Trap Weight, lbs.	1⅜	2	2¾	4	6	8⅝
Trap Length, ins.	2⅝	3	3⅜	3¾	4¼	4¾
Iron Str. List Prices	$3.30	$3.70	$4.40	$5.60	$7.20	$10.80
Iron Str. Weight, lbs.	2¼	3½	5	8	11	18¼
Iron Str. Length, ins.	3¾	4¾	5	5⅝	6¼	7½

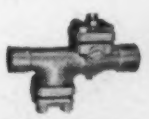
(Figure No. 140 to 147) Integral Strainer Impulse Trap for pressures up to 600 lb. Steel body, flanged connections.



(Figure No. 90 to 97) Impulse Trap for pressures up to 600 lb. (Using separate strainer.) Steel body, flanged connections.



(Figure No. 250 to 257) Integral Strainer Impulse Trap for pressures up to 600 lb. Steel body, socket weld connections.



(Figure No. 410 to 417) Integral Strainer Impulse Trap for pressures up to 150 lb. Bronse body, screwed connections.

THE YARWAY FINE-SCREEN STRAINER

Recommended for every Impulse Trap and for other pipe line appliances. Cadmium plated. Iron body for pressures up to 250 lbs., steel body from 250 to 600 lbs. Monel woven wire screen and easy-to-remove blow-off bushing. Six sizes, ½" to 2". Send for Bulletin S-201.

OTHER YARWAY PRODUCTS: Blow-off Valves for Power Plant Boilers · Blow-off Valves for Oil Field Boilers · Water Columns and Gauges · Remote Liquid Level Indicators and Recorders · Spray Nozzles · Expansion Joints

De Laval S-N Acid Treating Process

THE DE LAVAL SEPARATOR COMPANY

THE De Laval S-N continuous acid treating process as developed by The De Laval Separator Company, New York, and The Separator Nobel Company, Stockholm, Sweden, is applicable to the refining of the complete range of oils from cracked naphtha, pressure distillate, kerosene, etc., down to heavy cylinder stocks with viscosities as high as 200 S. S. U. at 210° F.

A feature of this process is the De Laval Acid Sludge Separator of which several types are available. As this centrifugal separator instantaneously and completely separates the acid sludges from the various oils and continuously discharges both oil and sludge, the time of contact between acid and oil may be carefully regulated to the optimum, rather than being determined by the length of time required for the sludge to settle by gravity.

While the problem of acid treating light oils is somewhat different from the refining of heavy lubricating stocks, with the exception of a few details of design, in general, the same equipment is used. The continuous process has been made feasible by the development of highly specialized centrifugal separators, and of almost equal importance, by the development of properly designed auxiliary equipment to effect the proper treatment of the oil and at the same time to condition it so that the sludge can be most efficiently removed by the centrifugal machine. As the acid and oil are carefully metered, continuously mixed and subjected to a predetermined degree of agitation, the acid is more efficiently utilized with a consequent reduction in acid requirements.

The proper control of contact time between acid and oil, in the case of cracked distillates, results in lower polymerization losses and lower loss of "anti-knock" rating during acid treatment; in the case of lubricating oil stocks, re-solution of the sludge is eliminated with its resultant darkening and instability of the oil. Batch agitators with air agitation with their inefficient use of chemicals, inflexible operation and difficult disposal of hard, viscous sludges, are eliminated.

Due to the more complete separation of the acid from the oil, in all cases there is a reduction in refining losses ranging from 25 to 50 percent of those encountered when gravity settling is employed.

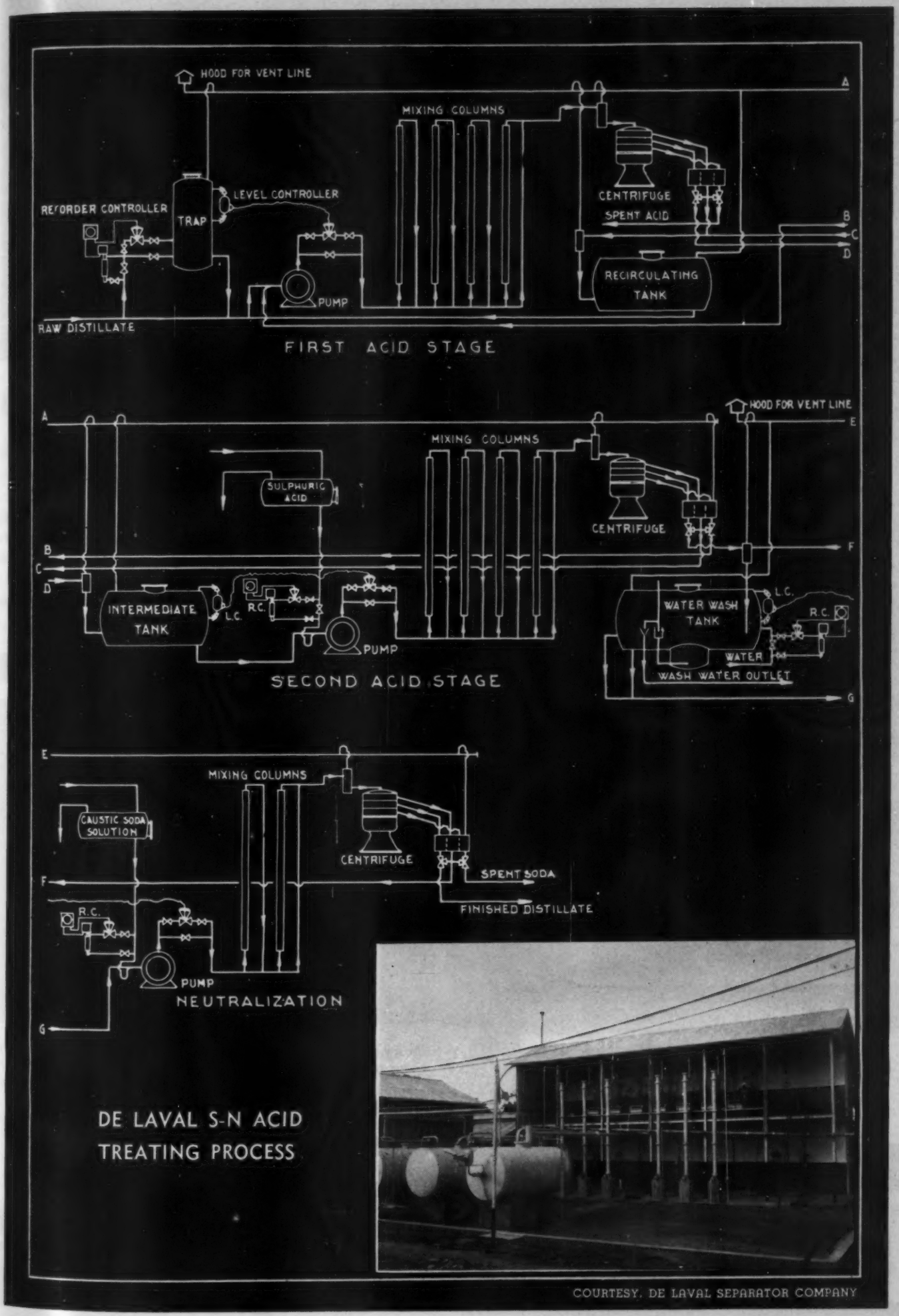
The flow sheet shows the equipment used in the treatment of light distillates. With a few modifications, this same flow diagram applies to the acid treatment of lube oils.

The operating sequence is as follows:

The raw distillate is metered through a recorder controller into a trap for water and sediment, which at the same time serves as a constant level suction tank for the acid mixing pump. The used acid from the second (fresh acid) separator flows into the suction of this pump. The mixture of distillate and acid is discharged either directly or through one or several of the contact columns to the first stage De Laval Acid Sludge Separator where complete separation takes place.

The spent acid sludge discharged from the centrifugal separator is less viscous and can be disposed of to waste more readily than that drawn from batch agitators. The pretreated distillate flows to an intermediate tank, from which is is picked up by a mixing pump. A controlled amount of fresh concentrated sulphuric acid is admitted into the suction of this pump. The mixture from this pump passes directly or through one or several of the mixing columns into the second stage De Laval Acid Sludge Separator. The acid treated distillate is discharged into an intermediate tank for continuous water washing with a controlled amount of water. The water-washed distillate is picked up by the mixing pump for neutralization. A controlled amount of caustic soda solution is introduced into the suction of this pump. The mixture is passed through the contact equipment into the De Laval Soda Separator. The neutralized distillate flows into a rundown tank, while the spent soda solution is discharged to waste. For convenience of operation during the starting and shutdown periods the plant is equipped with a small recirculation tank. The whole plant is completely enclosed and vapor-tight.

The plant is provided with special mechanical means for cleaning the separator bowls which usually takes place as a routine operation every second or third week.



COURTESY, DE LAVAL SEPARATOR COMPANY

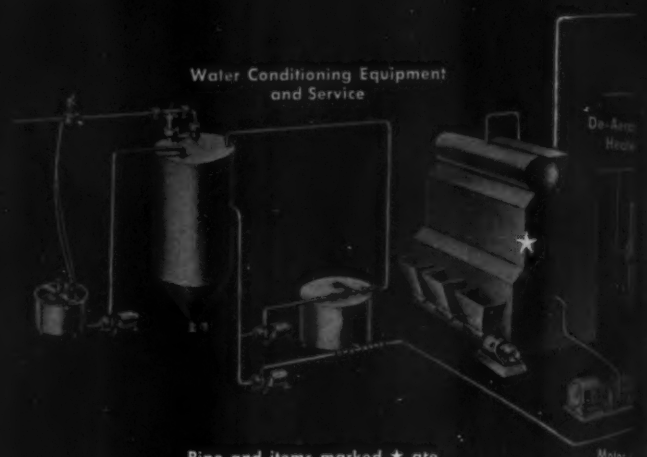
PETROLEUM EQUIPMENT "Buying Guide"

FOR FIELD, REFINERY, TRANSPORTATION

Need help for *new building — modernizing — expanding?* Get *practical, down-to-earth* buying assistance from Allis-Chalmers oil industry specialists, who *know* petroleum equipment needs!

And be *assured* of *unbiased recommendation* because you choose from the world's *widest range* of electric power and mechanical processing equipment!

A-C specialists are as near as your phone. Call your A-C Office or write ALLIS-CHALMERS, MILWAUKEE 1, WIS.

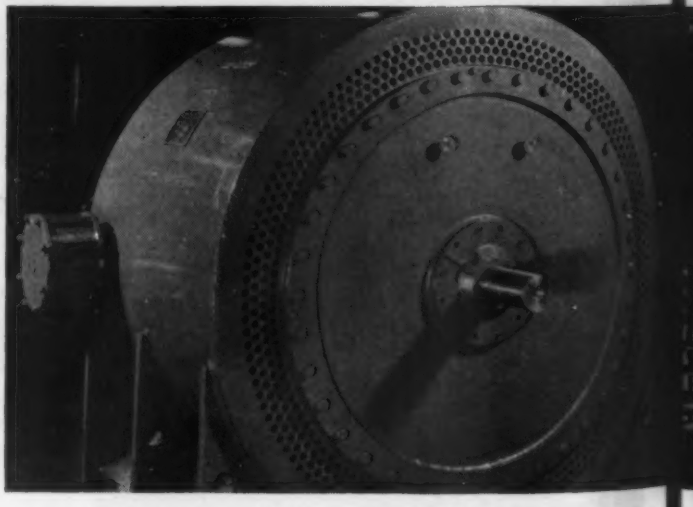


Pipe and items marked ★ are only items A-C does not build

Power Generation

THIS IS *YOUR* POWER SYSTEM . . . in one form or another. Note that all but two pieces of equipment can be supplied by Allis-Chalmers . . . *builders of the most complete line of major equipment for the generation and distribution of power.*

Each of these Allis-Chalmers products is available in a wide variety of types and sizes . . . standard or specifically engineered. This range of equipment simplifies your ordering problems . . . means you can

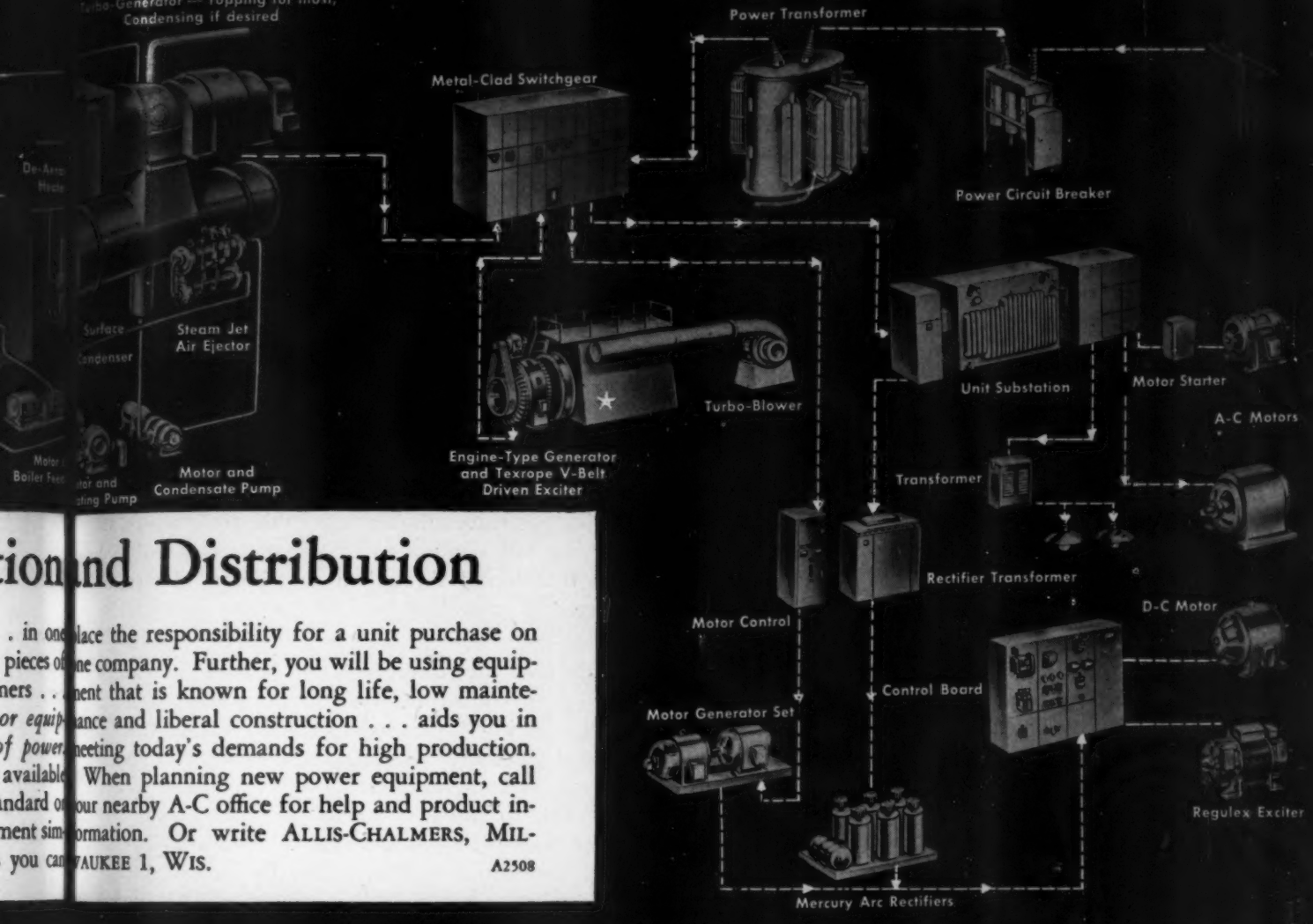


IN RATINGS TO 2,000 HP AND HIGHER, Allis-Chalmers builds these totally enclosed fan-cooled motors for hazardous locations . . . in refineries and pipelines. Simple ventilation system with tube-type air-to-air heat exchangers is easy to clean. Designed for service in corrosive atmospheres and in localities where air-borne particles create a problem in ventilating open-type motors.

Texrope and *Regulex* are Allis-Chalmers trademarks.

ALLIS

Power, Electrical, Proc Gen



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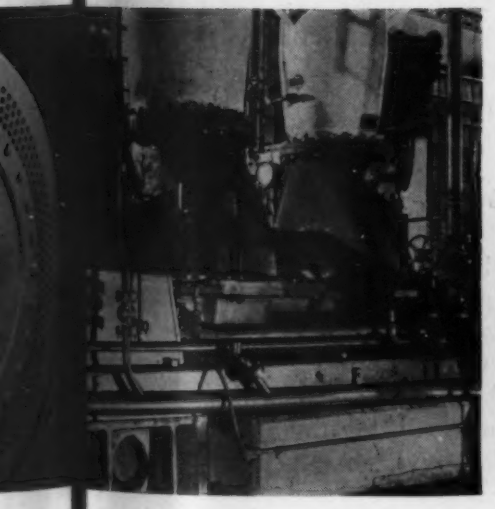
. . . in one place the responsibility for a unit purchase on pieces of one company. Further, you will be using equipners . . . ment that is known for long life, low maintenor equip- ance and liberal construction . . . aids you in f power meeting today's demands for high production. available When planning new power equipment, call ndard of our nearby A-C office for help and product inment sim- formation. Or write ALLIS-CHALMERS, MILyou can WAUKEE 1, WIS. A2508



CENTRIFUGAL BLOWERS . . . ROTARY COMPRESSORS . . . vacuum pumps are available in a wide range of types and sizes. Drives include motors, steam turbines, gas turbines—A-C is the *only* U. S. company to furnish gas turbine-axial compressor units for the Houdry process.

FOR PUMPING GASOLINE, WATER and other liquids, A-C can supply pumps in many types and sizes. Included is the Self-Priming unit, popular for loading and unloading tank cars, barges, etc. because it maintains pumping action even when slugs of air or gas are admitted.

SEND FOR FREE BULLETINS! The "Directory of Products" lists all Allis-Chalmers products and literature. Send for your copy . . . learn what A-C builds to help you maintain high production.

SCHALMERS

, Proc General Machinery for the Petroleum Industry

Alkylation Acid Regeneration Process

CHEMICAL CONSTRUCTION CORPORATION

THE CHEMICO Alkylation Acid Recovery Process is a development of Chemical Construction Corporation, New York. This method converts spent alkylation acid, regardless of its strength and carbon content, into new, water-white H_2SO_4 of any desired concentration. The system can also make additional acid from elemental sulphur or sulphur gases, either alone, or while spent acid is being reclaimed.

In brief, the process decomposes the spent alkylation-acid mixture by burning. The hydrocarbon content acts as fuel, and the acid itself breaks down into sulphur dioxide and water. The sulphur dioxide then is cooled, scrubbed, and passed into a contact acid plant, from which it emerges as new sulphuric acid, usually 98 percent H_2SO_4.

As shown in the diagram, the spent acid is pumped from a storage tank through an atomizing burner to the furnace, which is a brick-lined steel shell. The acid is decomposed at approximately 2000° F. with additional fuel or high oil sludges supplementing the hydrocarbons of the spent-acid mixture, if necessary. Preheated air is used for combustion. If desired, molten sulphur or hydrogen sulphide gas from other sources also can be burned in the furnace to supply heat and produce additional SO_2.

From the furnace, the gases pass to the recuperator, a tubular heat exchanger, and are cooled to 1200° F. The extracted heat is used to preheat the air fed into the furnace. From the recuperator, the gases go to the scrubbing tower. This tower is of steel, lined with sheet lead and acid-proof masonry, and is packed with spiral rings. Circulating dilute acid cools the gases to 250° F. and also washes out any dust which may have blown over from the furnace.

Next step is the cooling tower which is like the scrubbing tower, except it is not packed. The gases are cooled in this tower to 100° F. by water spray, and then go to the Cottrell mist precipitator. The SO_2 that dissolves in spray water of the cooling tower is removed in a stripping tower which is a lead-lined steel shell packed with spiral rings. A current of air passing through the water displaces the SO_2 gas which is returned to the main gas stream flowing to the precipitator.

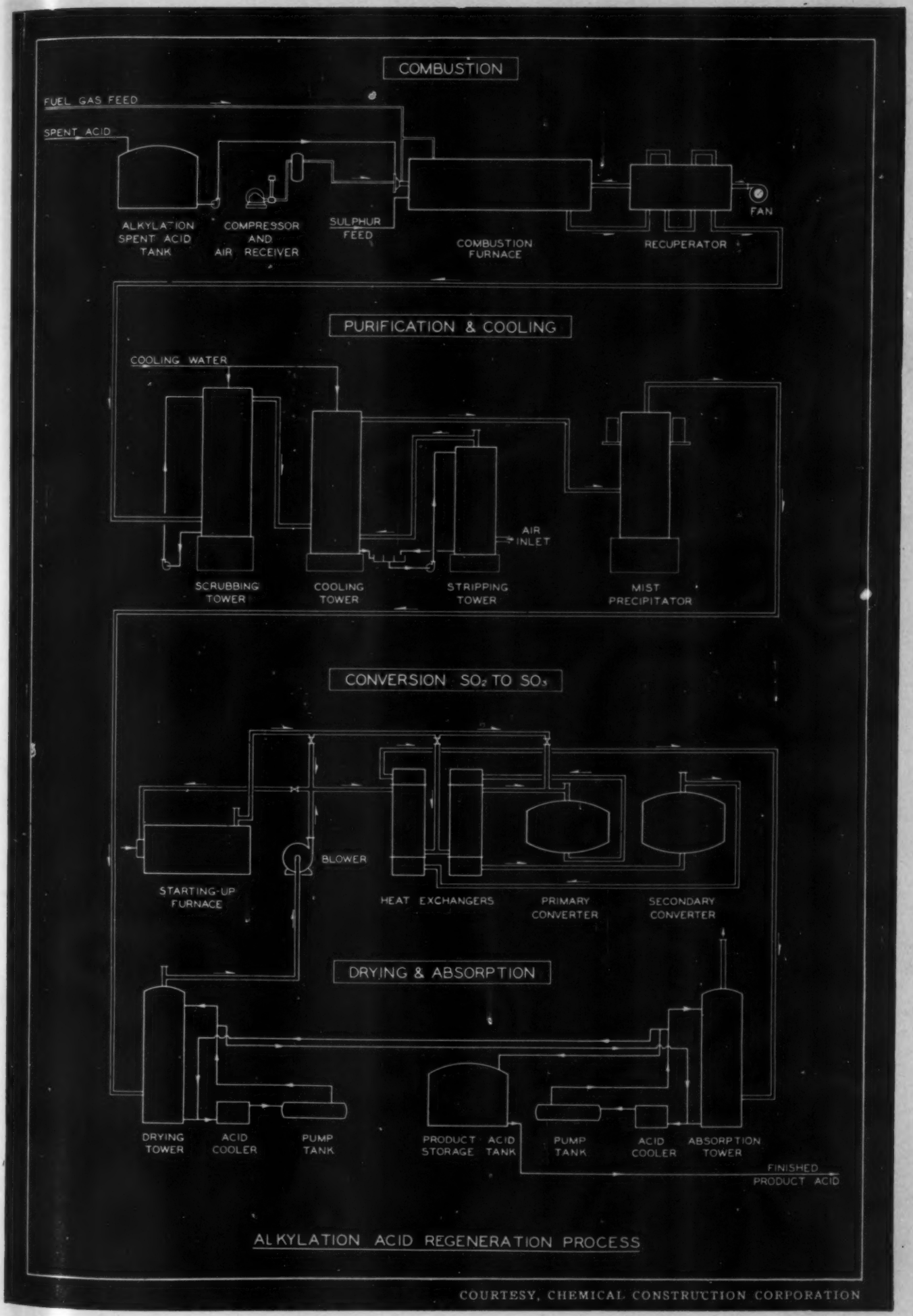
In the precipitator all the sulphuric acid mist in the gas stream is electrostatically removed and sent back to the scrubbing tower. The gas now is passed through a drying tower which has 93 percent sulphuric acid circulating over spiral ring packing. This acid is supplied from a pump tank and is cooled to desired temperature by cooling coils on its return to pump tank; sufficient 98 percent acid is constantly added to maintain the drying acid at 93 percent. As the dried gas leaves the drying tower air is added to reduce the SO_2 content to 8 percent, and the diluted gas is transferred through a centrifugal blower to primary and secondary heat exchangers for heating to 825° F.

Ready now for conversion, the heated gas enters the primary converter. This is an insulated steel shell containing layers of vanadium catalyst. The SO_2, in contact with the catalyst, is oxidized to SO_3, and rises in temperature to 1150° F. The hot gas is piped back to the hot side of secondary heat exchanger where it is cooled again to 825° F., the extracted heat being transferred to the incoming SO_2-and-air mixture entering the converter. The cooled gas goes now to the secondary converter and again is passed over vanadium catalysts to make sure all SO_2 is oxidized. This second oxidizing procedure generates more heat, which is removed by passing the gas through the hot side of the primary heat exchanger where the heat is absorbed by the SO_2 and air approaching the converters.

The SO_3 gas, at 250° F., now enters the absorption tower, a steel shell structure lined with acid-proof masonry and packed with spiral rings. The SO_3 is absorbed by the 98 percent acid circulating through the tower from the strong-acid pump tank and strong-acid cooling coils. Part of the 98-percent acid circulates back to the drying tower to maintain the drying acid at 93 percent and bring the absorbed water over to the absorption tower where it forms H_2SO_4 with the SO_3 gas being absorbed. Additional water is added as required.

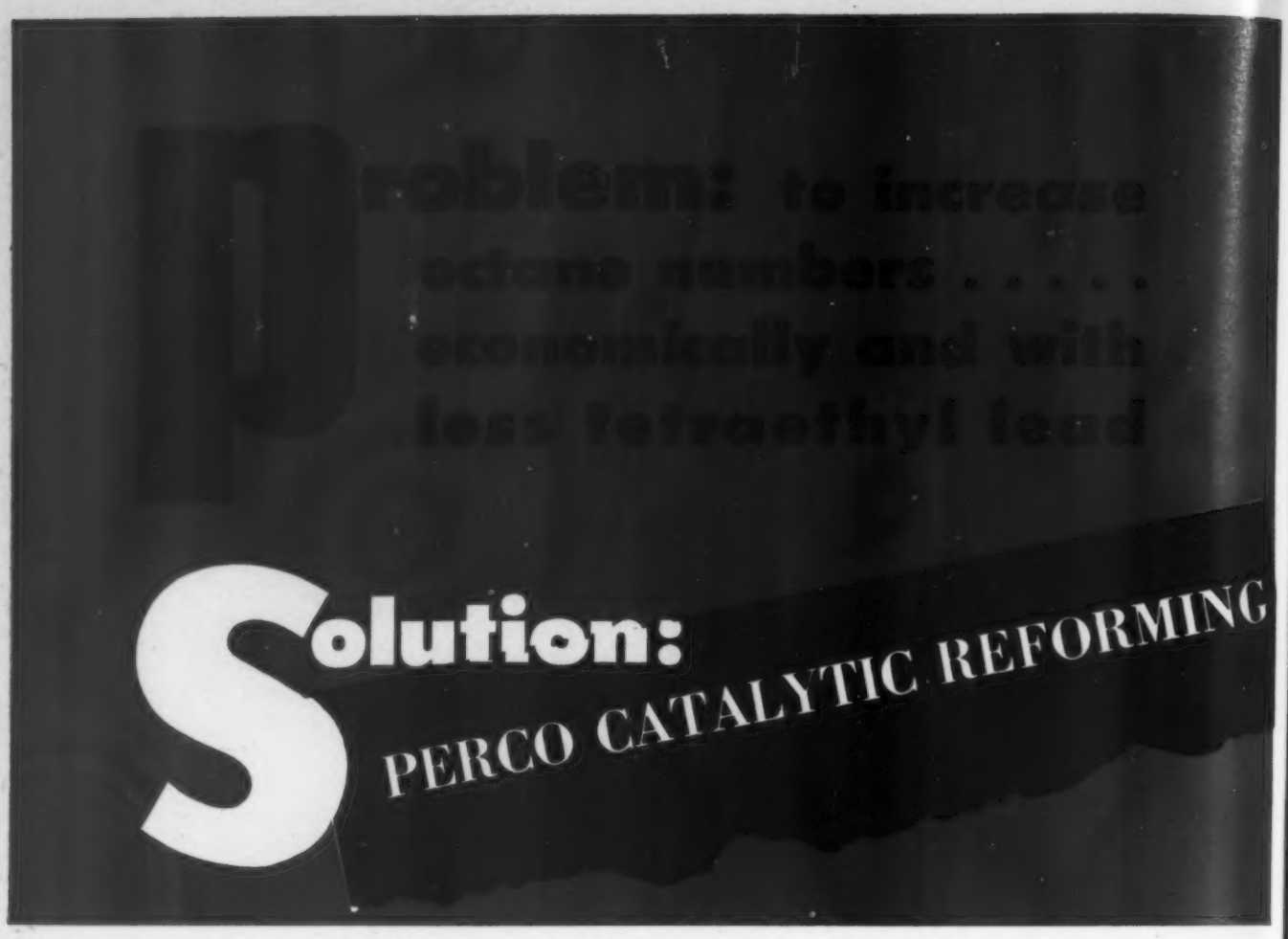
The storage tank receives the newly formed 98-percent acid as it is drawn off from the strong-acid cooling system.

In getting the system started, heat is supplied to the converters by an external starting-up furnace.



ALKYLATION ACID REGENERATION PROCESS

COURTESY, CHEMICAL CONSTRUCTION CORPORATION



PERCO Catalytic Reforming enables refiners to improve the octane numbers of their *total* refinery gasolines. The process treats many different feed stocks — straight run, cracked and polymerized gasolines, naphthas or blends — from a wide variety of crudes. By operating under mild catalytic cracking or catalytic reforming conditions all the advantages of desulfurization are obtained plus further appreciation in the ASTM octane number. The spread between ASTM and Research octanes is also increased.

Initial investment costs and operating expense of Perco Catalytic Reforming are low and the process is adaptable to any size refinery — large or small. A Perco engineer will gladly explain how this *modern* refinery method can be of benefit to you.

PHILLIPS PETROLEUM COMPANY

Chemical Products Department

BARTLESVILLE, OKLAHOMA

Progress with

PERCO PROCESSES

Gasoline and Naphtha Treating Processes

The continuing heavy demand for petroleum products coupled with the gradual depletion of low-sulfur-content crude oils in the United States is forcing refiners to process increasing amounts of high-sulfur-content crudes of which there are considerable reserves.

The problem of handling high-sulfur crudes has been worked out in two general ways by different refineries. In some cases refiners have built or converted from other service entire separate processing facilities for high sulfur crude operations. Others have elected to blend high-sulfur with low-sulfur crudes and to process the mixture. Wartime experience indicated that up to 5 percent of high-sulfur crude could, for instance, be blended with Midcontinent pipe line crudes without imposing any great hardship on the refineries served.

This general increase in crude oil sulfur content is especially noticeable in its effect on the product quality of raw gasolines and light distillates, and is greatly aggravating the treating problems encountered with these fractions.

The basic properties of gasolines that are adjustable by various treating methods include total sulfur content, corrosive sulfur content, color, color stability, gum content, and potential gum content. The extent and manner of treatment used to effect improvement in these properties also has a marked influence on lead susceptibility, antioxidant susceptibility, and octane number of the treated gasoline. With high octane gasolines in ever stronger demand, and with the general level of octane numbers going stadily upward, these secondary effects of treatment are receiving more attention as they have high economic importance.

Earlier treating methods involved use of sulfuric acid contacting, washing with aqueous alkali, and with alkaline plumbite. Sulfuric acid does effect substantial improvement in gasoline properties, but has a deleterious effect on octane number and lead susceptibilities. In general, its use has been abandoned as other methods have proved more efficient and less expensive. Thermally cracked gasolines undergo polymerization effects, too, when treated with sulfuric acid and require redistillation.

The once serious problem of gasoline color now is of little consequence because of the general adoption of dyed motor fuels. A degree of color improvement sufficient for present needs is usually difficult to obtain only where sweetening difficulties are encountered.

With high octane numbers for a major objective, the typical refinery now employs a variety of treating processes ranging from simple caustic washing through various sweetening processes to mild catalytic reforming.

Solutizer Process

SHELL DEVELOPMENT COMPANY

THE Shell Solutizer Process for sweetening gasolines and naphthas is unique in that a doctor-sweet product is made by substantially complete physical extraction, instead of chemical conversion, of mercaptans. For the extraction, caustic solution containing an organic substance (such as potassium isobutyrate), called for brevity Solutizer, is used. The removal of mercaptans gives products characterized without exception by an improvement in their TEL susceptibility and, very often, by an improved initial octane rating, compared to those sweetened by conventional doctor or other treatments based on the oxidation of mercaptans to disulfides. In addition, there is a reduction in total sulfur brought about by the Shell Solutizer sweetening which in some cases is sufficient to eliminate sulfuric acid treatment normally required to produce fractions conforming to low-sulfur specifications. In other cases, the Shell Solutizer sweetening applied prior to acid treating results in the reduction of acid requirements. However, even when the Shell Solutizer Process replaces doctor or other sweetening treatment of acid-treated stocks, substantial economies and improvements in product qualities are effected.

The process is applicable for sweetening either straight-run cracked, reformed, or synthetic fractions, and its effectiveness even extends to the products boiling outside the normal gasoline range.

The Shell Solutizer Process is continuous and fully regenerative. Essentially, it consists of mercaptan extraction with a lean solutizer-solution and the regeneration of the fat solutizer-solution by boiling and stripping with steam. As indicated on the accompanying flow diagram, a sour gasoline or other hydrocarbon fraction is first pre-treated with a caustic solution of suitable concentration to remove such impurities as carbonoxylic and other relatively strong acids or to adjust the concentration of some acidic components, such as alkyl phenols, to permit efficient solutizer treatment. Usually a batch of caustic solution is used for pre-treating until practically exhausted. The pretreated gasoline is passed on to a countercurrent extraction unit in which it is brought into contact with a lean solutizer-solution. Extraction columns packed with carbon Raschig rings are effective and proven for this purpose. The sour hydrocarbon fraction is introduced near the bottom of the column and the lean solutizer-solution enters near the top of the column. Intimate countercurrent contact of the two streams results, and a doctor-sweet fraction and fat solutizer-solution are withdrawn from the opposite ends of the column.

The fat solutizer-solution then is diluted with a condensate produced in the regenerating step. The diluted solution is allowed to settle in a separator, where a small quantity of sour gasoline, initially partly entrained and partly dissolved in the fat solution, separates out as the result of the dilution. This sour gasoline is skimmed and returned to the extraction column. The separated fat solutizer-solution is preheated and fed to the top plate of a bubble-plate solutizer-solution regenerator. The fat solution is regenerated by being stripped of mercaptans by steam either injected from an outside source or generated in a reboiler at the bottom of the regenerator. A temperature in the neighborhood of 270° F. is maintained at the bottom of the column; it varies somewhat with the composition of the solutizer solution and the pressure. The regenerated or lean solution is pumped through heat exchangers and coolers to the top of the extraction column.

The steam and mercaptans pass from the regenerator to the condensate stripper and on to, first, the condenser and, then, to a separator. The condensate is pumped from the latter to the top of the condensate stripper, from which a mercaptan-free condensate is withdrawn to dilute the fat solutizer-solution just prior to its entering the gasoline separator. The mercaptans separated from the condensate are available for chemical syntheses or can be disposed of as fuel.

The construction of the Shell Solutizer Process plant presents no unusual engineering or corrosion problems. With the exception of a copper-nickel-alloy tube-bundle in the reboiler, the equipment can be made of low-carbon steel.

One of the developments of the basic process permits, under certain circumstances, a treatment of several gasolines to produce their doctor-sweet blend at practically no additional cost over that of the solutizer sweetening of one of the component gasolines.

This process is licensed by Shell Development Company, New York.

SOLUTIZER PROCESS
COURTESY, SHELL DEVELOPMENT COMPANY
Doctor Sweet Gasoline
10000
PACKED COLUMN
100
35
Lean Solutizer Solution
Condensate
20000
58.5
225
5
CONDENSATE STRIPPER
Reflux
CONDENSER
Mercaptans
Stm
MERCAPTAN SEPARATOR
REBOILER
SOLUTIZER REGENERATOR
9
270
Fat Solution
HEAT EXCHANGER
GASOLINE SEPARATOR
Fat Solution
SOLUTIZER EXTRACTOR
100
50
CAUSTIC PRETREATER
5% NaOH
MIXER
65
CAUSTIC SETTLER
Sour Gasoline
10000
585
20000
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE DEGREES FAHRENHEIT

Mercapsol Gasoline Treating Process

THE PURE OIL COMPANY

THE Mercapsol Treating Process is a practical, economical method for removing mercaptans from gasoline. In this process the removal of mercaptans is accomplished by extraction with an aqueous solution of caustic soda containing certain solubility promoters. Continuous, countercurrent extraction is performed in a packed tower, and the spent treating solution is continuously stripped of mercaptans by steam in a bubble tower or packed tower. Provisions are made for protecting the treating solution against the accumulation of sodium sulfide and other undesirable substances. Auxiliaries are included for removing all traces of treating solution from the treated gasoline and for recovering entrained and dissolved gasoline from the spent treating solution.

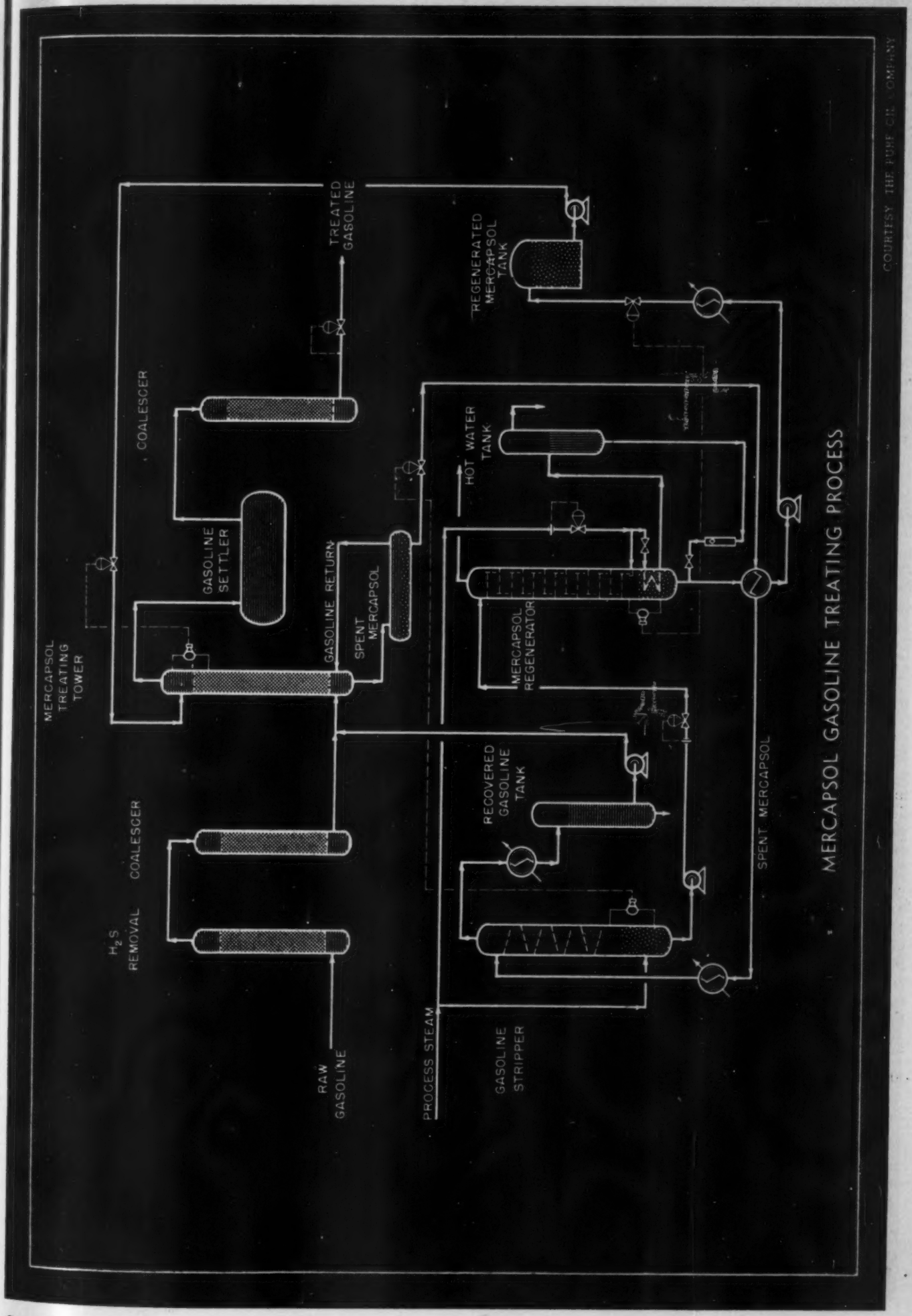
The equipment used for the Mercapsol process is all of conventional type. In fact, many items of existing plant equipment can usually be incorporated into a Mercapsol unit, and multi-stage mixing and settling can be substituted for packed tower extraction. The treating solution is made from inexpensive materials and can be used indefinitely, only a small amount of chemical makeup being necessary because of leakage and other minor losses. The operation of the process is simple and can be made practically automatic by the use of suitable instruments.

In the operation of the process, raw gasoline is fed to the H_2S removal tower (packed) and flows upward through a batch of caustic soda solution. The gasoline then passes down through a packed coalescer whose function is to remove the aqueous phase and anything contained therein, from the gasoline. Then the gasoline enters the mercaptan removal tower and flows through the packing countercurrently to the down-coming stream of Mercapsol solution, which extracts substantially all of the mercaptans from the gasoline. The gasoline then goes to a settler and from there through a coalescer for removal and recovery of the last traces of treating solution before going to storage. A pressure regulator on the gasoline run-down line holds sufficient back pressure on the system.

A stream of regenerated Mercapsol is pumped from the storage tank into the mercaptan removal tower near the top and is regulated by a liquid level controller operating on the interface between gasoline and Mercapsol in this tower. Mercapsol solution flows from the bottom of the mercaptan removal tower to a settler in which entrained gasoline is separated and returned to the tower. Then the spent Mercapsol exchanges heat with the hot regenerated solution in an exchanger and is further heated by steam in a heater before going to the top of the gasoline stripper. The flow of Mercapsol into this stripper is regulated by a level controller near the bottom of this tower. A small amount of process steam flowing up through the stripper removes the small amount of dissolved gasoline, which is condensed, collected in a small tank, and pumped back to the mercaptan removal tower. This stripper is designed and controlled to remove substantially all the gasoline from the Mercapsol without removing more than a very small portion of the mercaptans present in the solution. Mercapsol, at a rate set by a recording flow controller, is pumped from the bottom of the gasoline stripper to the top of the regenerating tower. Flowing down counter-currently to the steam rising in this tower, the Mercapsol is stripped of mercaptans. A liquid level controller holds a level in the bottom of the regenerator, and regenerated Mercapsol, after passing through the heat exchanger, is pumped through a cooler to the storage tank. Stripping steam is regulated by a recording flow controller, and a portion of this steam is used in a closed coil to keep the Mercapsol in the bottom of the tower boiling moderately. Steam condensate from this coil and from the steam heater mentioned previously is collected in a small tank, and, since this water is free of dissolved solids, it is used to replace the small amount of water evaporated from the Mercapsol during regeneration. The circulation rate of Mercapsol and the amount of stripping steam required depend upon the concentration and distribution of mercaptans in the raw gasoline.

The recent trend toward the processing of high-sulfur crudes has increased the attractiveness of gasoline treating methods which remove mercaptans over those which convert mercaptans to disulfides. Regenerative caustic scrubbing, which has been employed frequently in conjunction with the older treating methods, usually did not remove much more than half the mercaptan sulfur in the raw gasoline. Between 95 percent and 99 percent removal of mercaptan sulfur is obtained by use of the Mercapsol process, and this substantially greater reduction in sulfur results in considerable savings of tetraethyl lead as well as other benefits. In many cases the Mercapsol treated gasoline is "doctor-sweet," and in almost any case contains such a low concentration of mercaptan sulfur that its effect upon odor or lead susceptibility is not detectable.

The Mercapsol Process, in successful commercial operation since 1941, is licensed by The Pure Oil Company.


MERCAPSOL GASOLINE TREATING PROCESS
COURTESY THE LUMMUS COMPANY
TREATED GASOLINE
REGENERATED MERCAPSOL TANK
COALESCER
HOT WATER TANK
GASOLINE SETTLER
GASOLINE RETURN
SPENT MERCAPSOL
MERCAPSOL TREATING TOWER
MERCAPSOL REGENERATOR
RECOVERED GASOLINE TANK
SPENT MERCAPSOL
H_2S REMOVAL
COALESCER
RAW GASOLINE
PROCESS STEAM
GASOLINE STRIPPER

Cobalt Molybdate Desulfurization Process

UNION OIL COMPANY OF CALIFORNIA

THE depletion of low-sulfur-content crudes in the United States is forcing refiners to employ increasing amounts of high-sulfur-content, low-quality crudes of which there are considerable reserves. To date, the utilization of these latter stocks has been largely in the form of blends with the sweet crudes normally used in refinery operations. The sulfur-containing distillates have been handled by a variety of methods ranging from mercaptan removal only to mild catalytic reforming. However as the amount of high-sulfur stocks processed increases, the proportion of refractory thiophenes and thiophanes, largely unaffected by present treating processes, also is increased, and a definite need is arising for a more effective means of desulfurization.

The Cobalt Molybdate Desulfurization Process has been developed as an answer to the limitations of existing procedures. It represents a versatile method of sulfur removal capable of handling all types of petroleum distillate stocks encountered in normal refining operations, and is capable of removing sulfur from even the most refractory compounds associated with these stocks. Fundamentally, the process involves the conversion of organically combined sulfur to hydrogen sulfide by contact with a catalyst containing cobalt molybdate in the presence of hydrogen. Cracked gasoline stocks containing 3 percent by weight or more of sulfur have been treated to give products of as low as 0.01 percent by weight of sulfur in yields of 99-100 volume-percent. Heavy gas-oil stocks containing as much as 5.0 percent by weight of sulfur have been reduced in sulfur content to less than 0.2 percent with 100 volume percent yields. The accompanying diagram presents a simplified process flow for a unit to handle coke-still distillate of about 4.0 percent sulfur content and containing approximately 30 percent of material distilling in the lubricating-oil range. Referring to the diagram, the feed stock and hydrogen are preheated by exchange against the hot reaction products. The exothermic heat of reaction is of sufficient magnitude to make the process independent of the preheat furnace except during start-up periods. The catalytic treatment may be conducted over a wide range of temperatures, but for maximum yield and desulfurization, a temperature in the neighborhood of 800° F. is employed. The reaction products after exchange with the feed are cooled and passed to a separator. The liquid product from the separator may be caustic-washed to remove traces of dissolved hydrogen sulfide and then sent to storage. The gas from the separator, containing hydrogen, hydrogen sulfide and small amounts of hydrocarbons, is treated in an oil absorption unit at the reaction pressure and the separated hydrogen recycled to the reactors.

For the treatment of coke-still distillate, a multiple reactor construction is employed, three reactors being on stream while the fourth is being regenerated. Regeneration of the catalyst is accomplished in the conventional manner by adding air to a stream of inert combustion gas circulated through the catalyst bed. The characteristics of a typical coke-still distillate and the products resulting from its treatment by the Cobalt Molybdate Process are given in the following tabulation:

Characteristics of Feed and Products from Desulfurization of Coke Still Distillate

	Feed, Heavy Coker Gas Oil	Gasoline Fraction of Product	Gas Oil Fraction of Product
Yield, percent by Vol. based on Feed	....	8.4	91.1
Gravity, °API at 60° F.	19.4	47.7	28.1
ASTM Distillation Temp., °F.:			
Initial	482	164	480
10 percent	584	238	498
50 percent	699	320	612
90 percent	....	372	....
Maximum	760+	392	760+
Rec. percent	72.5	98.0	91.0
Sulfur Content, percent by weight	4.2	0.1	0.5
Acid Solubility, percent by Vol.	100.0	37.5	50.0
Knock Rating, CFR Motor Method:			
Clear	....	64.5	....
3.0 ml. TEL/gal.	....	76.5	....
Viscosity, SUS at 100° F.	84	....	55
Pour Point, °F.	65	....	45

The above is illustrative of only one type of application of the Cobalt Molybdate Process. Cobalt Molybdate treatment is unusually flexible and, if desired, can be used not only to remove sulfur compounds but simultaneously to improve the octane rating of gasoline stocks. In all cases the gasoline products have excellent lead susceptibilities. The products heavier than gasoline make good reforming and cracking stocks.

Detailed information on specific problems can be obtained from Union Oil Company of California, Los Angeles.

COBALT MOLYBDATE DESULFURIZATION PROCESS
COURTESY, UNION OIL COMPANY OF CALIFORNIA
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
REACTOR FURNACE
INERT GAS GENERATOR
REACTORS
GAS VENT
GAS SEPARATOR
CAUSTIC SCRUBBER
WATER WASHER
OIL ABSORBER
STRIPPER
FUEL GAS
AIR
FEED STOCK – 3.9 % S
HYDROGEN
CAUSTIC
WATER
DESULFURIZED PRODUCT
0.4 % S
30° API
15 040
25° API
9 000 000
H_2S
700
1000
700
700
800
450
100
200
425
100
100
400
50
6600

Question:

What treating process can add "new refinery" economies in your modernization plans?

Answer:

The TANNIN SOLUTIZER PROCESS . . .

Here's why:

You immediately achieve substantial savings in TEL.

You profit through reduced loss of gasoline.

Your operating costs are extremely low with this thoroughly modern yet simple process.

With the Tannin Solutizer Process, you can generally adapt your existing treating equipment with almost negligible conversion costs.

We will be glad to give you more information on this established process. The TANNIN SOLUTIZER PROCESS is licensed under the rights of Socony-Vacuum Oil Company and Shell Development Company by

SHELL DEVELOPMENT COMPANY, INCORPORATED

50 West 50th Street, New York 20, N.Y.

SHELL

Why Chemico-built Plants are Profitable

Improved Processes

A chemical plant is only as good as the process it uses. Chemico originates, controls, and provides many of the most advanced processes and equipment . . . develops entirely new processes where necessary.

Individually Designed

Chemico-built plants are designed to meet each client's requirements. Capacity, type of raw material, available utilities, location, space limitations . . . all are considered.

Economical Use of Manpower

Automatic operation, centralized control, efficient plant layout . . . reduce manpower requirements and make more productive use of every man-hour.

Minimum Maintenance

Maintenance problems are simplified through the use of carefully selected equipment and improved processes, and careful layout of the plant. Repairs are never "hard to get at."

Experience

Chemico designs and builds many kinds of heavy-chemical plants . . . has been doing this for 34 years. Over 600 installations the world over are evidence of Chemico's successful experience.

Chemico furnishes these plants under one contract—one responsibility—with an iron-clad performance guarantee . . . is prepared to undertake other similar projects on the same basis.

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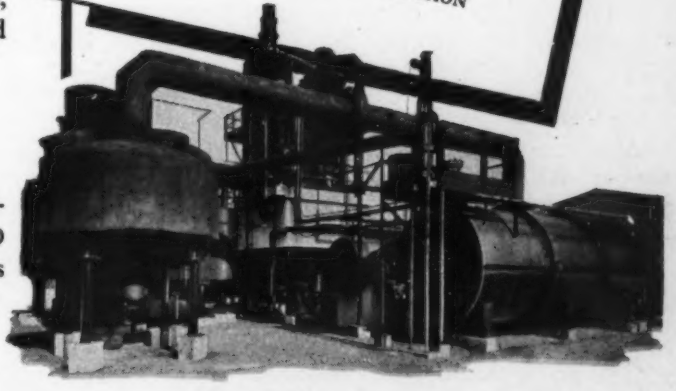
N. E. C. SYNTHETIC METHANOL
NITRIC ACID
PHOSPHORIC ACID
COMPLETE FERTILIZER PLANTS
SULFURIC ACID
PRODUCTION, CONCENTRATION, RECOVERY
PETROLEUM SLUDGE CONVERSION
FORMALIN
WILTON PROCESS TAR DISTILLATION



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EUROPEAN TECHNICAL REPRESENTATIVE
CYANAMID PRODUCTS, LTD., BRETTENHAM HOUSE, LANCASTER PLACE, LONDON W. C. 2, ENGLAND
EUROPEAN LICENSEE OF N. E. C. PROCESS
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CABLES: CHEMICONST, NEW YORK

"Chemico plants are profitable investments"

CC 103

Tannin-Solutizer Process

SHELL DEVELOPMENT COMPANY

MERCAPTAN removal processes in general may be classified into two types: those that effect sweetening by conversion of the mercaptans to other sulfur compounds, usually disulfides, and thereby result in little, if any, reduction in sulfur content; and those that achieve a physical or chemical removal of the mercaptans and substantially an equivalent reduction in the total sulfur content of the hydrocarbon treated. The Tannin-Solutizer Process is a member of this latter group, and thus obtains the product improvement and increase in octane number and lead susceptibility to be derived from sulfur reduction. The process obtains this objective by addition of a simple and efficient tannin-catalyzed oxidation regeneration step to a mercaptan extraction operation employing strong caustic solutions containing mercaptan solubility promoters, termed "solutizers," such as isobutyric acid and/or alkyl phenols.

The process, as illustrated in the flow diagram, consists essentially of three sections: the caustic pretreat, solutizer treatment, and regeneration system. Caustic pretreatment usually is required for removal of strongly acidic components, such as hydrogen sulfide, and in some cases for control of alkyl phenol concentration. A single conventional mixer-settler contact stage is normally satisfactory for this purpose, and the caustic solution may be used until essentially saturated.

After pretreatment the sour hydrocarbon feed enters the extraction system in which it is countercurrently contacted with a suitable solutizer solution. The solution may contain added isobutyric acid solutizer, or naturally occurring alkyl phenols may be utilized for this purpose if their concentration is sufficient in the hydrocarbon fraction to be treated to maintain the required solutizer strength in the treating solution. Since only a nominal concentration of alkyl phenols in the gasoline is necessary for this purpose, most plants are self-sufficient and solution costs are exceedingly low. The extraction equipment may be composed of mixer-settler stages, such as the typical three-stage system shown on the flow diagram, or other suitable contacting equipment such as towers packed with Raschig rings or of perforated tray design. The sweetened hydrocarbon flows from the extraction system to storage or subsequent operations.

Treating solution from the extraction step is preheated to a temperature of 110-130° F. and flows to the regeneration column in which the extracted mercaptans are catalytically oxidized to the corresponding disulfides by air blowing. This operation may be accomplished in a column devoid of packing equipped with efficient air dispersers or in any standard liquid-gas mixing device. The regenerated solution is coalesced for removal of entrained disulfide oil and the portion of this material remaining in solution is removed by a simple countercurrent or batch naphtha-wash step. The naphtha requirements for virtual elimination of disulfides from the lean treating solution are small, amounting in volume to only 1 to 2 percent of the hydrocarbon-treating rate, and may be disposed of in third-grade fuel or added to reforming or cracking stock for conversion of the disulfides to hydrogen sulfide. The lean solution then is cooled and returned to the extraction system, thus completing the cycle.

Additional equipment consists of coalescers appropriately located to safeguard against entrainment of hydrocarbon or solution and a filter to prevent accumulation of possible side products of oxidation and extraneous material in the treating solution.

This process easily achieves mercaptan removal to the "odor sweet" 0.001 to 0.003 weight-percent mercaptan sulfur level when treating a wide range of straight-run or cracked stocks, either low-end-point or full-boiling-range gasolines, having a naphthenic or paraffinic crude source. Mercaptan reduction to the doctor-negative specification may be obtained in many cases by means of more stringent operating conditions and increased number of extraction stages. The process is characterized by a simple and flexible design which is particularly readily adaptable to utilization of existing equipment; and, due to the low operating temperatures and non-corrosive nature of the treating solution, no corrosion-resistant metals are required. Simplicity of operation and adaptability to virtually complete automatic control results in low labor costs, and the inexpensive solution has indefinite treating life so that chemical cost is nominal. The limited heating requirements may be fulfilled by exhaust steam. The profitability gained from the saving in TEL resulting from the sulfur removal in most cases exceeds this low overall operating cost, so that an early payout of the original capital investment of an installation is realized.

The Tannin-Solutizer Process is licensed by Shell Development Company under patents of Socony-Vacuum Oil Company and Shell Development Company.

TANNIN-SOLUTIZER PROCESS
SOLUTIZER TREATMENT
CAUSTIC COALESCER
CAUSTIC PRETREAT
1ST. STAGE MIXER-SETTLER
2ND STAGE MIXER-SETTLER
3RD STAGE MIXER-SETTLER
SOLUTION COALESCER
Sweetened Gasoline
Sour Gasoline
NaOH
REGENERATION SYSTEM
Exhaust Air
Fat Solution
Lean Solution
COOLER
FILTER
NAPHTHA WASH TOWER
SETTLING & SURGE TANK
HEATER
GASOLINE COALESCER
REGENERATION COLUMN
Air 40000
Wash Naphtha
RSSR + Naphtha
RSSR COALESCER
* Basis .04% w Mercaptan Sulfur removal
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
COURTESY SHELL DEVELOPMENT COMPANY

Unisol Mercaptan Extraction Process

UNIVERSAL OIL PRODUCTS COMPANY

FOR many years petroleum process development touched only lightly on gasoline treating methods. Within the last few years demand for higher octane numbers and increasing use of high sulfur crudes has focused attention on this problem. Removal of sulfur compounds from gasoline gives marked improvement in tetraethyl lead susceptibility, although the unleaded octane number may improve only slightly. Conventional gasoline-sweetening processes convert malodorous mercaptan sulfur compounds to corresponding alkyl disulfides without reducing sulfur content.

The Unisol process was developed to extract mercaptan compounds from gasoline, thereby accomplishing the purpose of conventional sweetening and at the same time improving tetraethyl lead susceptibility by removal of the sulfur associated with the mercaptan. Conventional sweetening never improves octane number and often reduces it, particularly with high mercaptan stocks, making sweetening a nonprofitable operation. Unisol, on the other hand, generally shows a rapid return on the investment through improved product quality and tetraethyl lead savings. The process is the culmination of years of large-scale pilot-plant development work by The Atlantic Refining Company. It is licensed by Universal Oil Products Company.

The Unisol process is based upon the large increase in solubility of mercaptans in caustic soda solutions when an organic solvent, methyl alcohol, is present.

Hydrogen sulfide-free gasoline is introduced to the bottom of an extraction column and has mercaptan extracted from it as it flows upward in countercurrent contact with an aqueous solution of methanol and caustic soda. The caustic solution is introduced at the top of the extractor and methanol near the mid-point. Above the methanol inlet the caustic solution extracts any methanol dissolved by the gasoline from the caustic - methanol solution. Treated gasoline is removed from the top of the extractor and sent to storage without any subsequent treating except addition of inhibitor. The fat caustic-methanol solution containing extracted mercaptan is removed from the bottom of the extraction column from whence it is charged to a column in which the methanol and mercaptan are stripped from the caustic solution. Regenerated caustic solution is recovered from the bottom of this column for reuse. From the top of the stripping column, stream-methanol-mercaptan vapors go to a condenser and then to a mercaptan separator from which the extracted mercaptans are decanted from the system. Aqueous methanol solution is charged from the mercaptan separator to a fractionating column from which methanol is recovered for reuse as an overhead product.

The process is outstanding in its simplicity, low cost and effectiveness. Mercaptan reduction of better than 99 percent is readily accomplished. Commercial units are operating profitably on gasolines containing as low as 0.013 and as high as 0.25 percent mercaptan sulfur. Octane number improvements at the 3-cc-tetraethyl-lead level, range from 1 to 3 or more on the commercial units, depending on mercaptan-sulfur content of the untreated gasoline.

The treated product as it comes from the unit usually is sweet or nearly sweet. In almost all cases the product is completely sweet after a few days in storage.

The process is noteworthy in its ability to control the gasoline-phenol content. Phenols can be left in the gasoline or can be almost completely removed as a concentrated sodium phenolate solution containing substantially no free sodium hydroxide.

In comparisons of the properties of Unisol-treated gasoline with the raw gasoline or with the product obtained by conventional sweetening, it generally is found that Unisol treating improves inhibitor susceptibility, lowers ASTM gum, and that copper-dish gums after inhibiting are considerably lower.

The ability of the process to extract mercaptans with a circulation of only 3-5 percent caustic-methanol keeps chemical consumption, pumping and utility costs for regeneration of reagents low. Combined chemical and utility costs average about 1 to 1½ cents per barrel treated for most gasolines.

The possibility of a return on the investment usually makes Unisol treating more attractive to install than a conventional sweetening unit. Many gasolines contain more mercaptan than a conventional sweetening unit will successfully handle, making it necessary to remove part of the mercaptan by regenerative caustic extraction. In these cases a Unisol unit will cost no more to install than the combined caustic extraction plus sweetening units.

Many Unisol units ranging in designed capacity from 500 to 25,000 barrels per day now are operating and amply paying their way in improved product quality and reduced tetraethyl lead requirements.

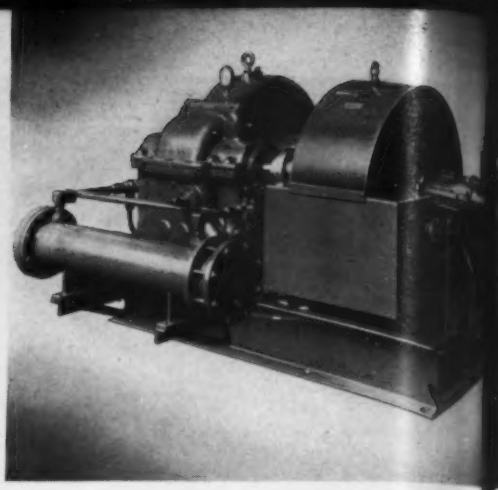
UNISOL MERCAPTAN EXTRACTION PROCESS
COURTESY UNIVERSAL OIL PRODUCTS COMPANY
METHANOL RECEIVER
METHANOL FRACTIONATING COLUMN
STEAM
WATER
MERCAPTAN
MERCAPTAN SEPARATOR
STRIPPING COLUMN
STEAM
METHANOL
CAUSTIC SOLUTION
TREATED GASOLINE
PACKED EXTRACTION COLUMN
PACKED H_2S SCRUBBER
SOUR GASOLINE

MOTORS

Elliott Fabri-Steel all-welded construction — split-sleeve bearings — uniform air gap — directed ventilation. Two-pole, for pipe lines and general service.

GENERATORS

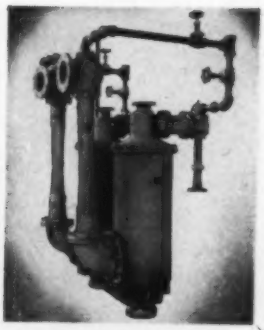
Elliott Fabri-Steel construction, with refinements in details, insure maximum performance in teaming up with gas or Diesel engines.



TURBINES

Interchangeable, corrosion-resistant parts — positive emergency speed control — spray-coated journals—center-line bearing supports —vital parts sealed—five sizes—eight types.

Where reliability and e
are valued-



EJECTORS

Elliott steam jet ejectors, with barometric condensers, for economically maintained vacuum on large refinery units.

STRAINERS

For water and oil service. Twin strainers, single strainers, self-cleaning water strainers, motor-driven.

ELLIOTT COMPANY JEANNETTE, PA.

BLOWERS

Multi-stage, single-stage, turbine or motor driven. For non-stop operation as gas boosters — for cat crackers—absorption processes—sulphuric acid plants — asphalt plants — oxygen and synthesis plants.

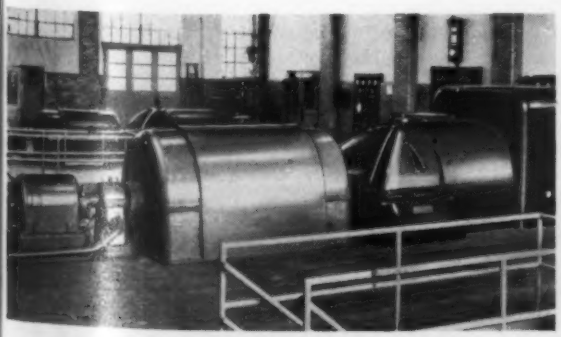
TUBE CLEANERS

Tube cleaning equipment that turns "Down time" into "Quick time", for every tube or pipe that must be cleaned — backed by 50 years' experience in solving tube cleaning problems.

TURBOCHARGERS

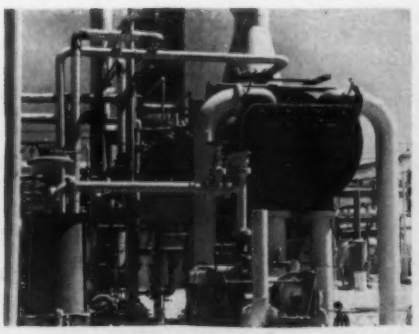
Elliott-Buchi turbochargers are everywhere boosting Diesel power, and saving fuel. Exhaust gases provide power—no additional engine power required—fully automatic. For four-cycle Diesel engines, 250 hp and larger.

economy it's Elliott



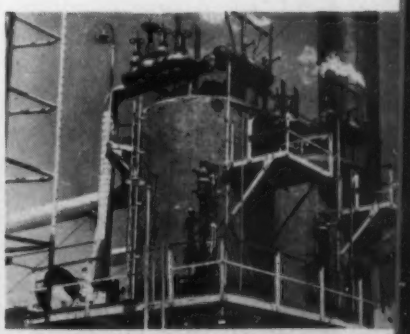
TURBINE-GENERATORS

Both turbine and generator designed and built together — a complete Elliott unit.



CONDENSERS

Surface and barometric condensers in capacities as required in oil industry power plants.



DEAERATING HEATER

Applying principles developed by Elliott in pioneering deaeration, delivering oxygen-free boiler feed water.

Q-1136A

RIDGWAY, PA. • SPRINGFIELD, O. • NEWARK, N. J. • DISTRICT OFFICES IN PRINCIPAL CITIES

Copper Sweetening Process

PHILLIPS PETROLEUM COMPANY

VARIOUS methods of treating gasolines with reagents comprising compounds of copper have been suggested for many years. Phillips Petroleum Company, through licensing activities of the Perco Division, Bartlesville, Oklahoma, has made available copper sweetening processes used by over 100 refineries and gasoline plants in the United States, and units are in use in Canada, South America, Europe and Africa.

These processes take advantage of the oxidation of mercaptans by cupric chloride, and the subsequent regeneration of the cuprous chloride with oxygen as shown by the equations:

(1) $2\ RSH$ (Mercaptan) $+$ $2\ CuCl_2$ (Cupric Chloride) $\rightarrow$ $RSSR$ (Disulfide) $+$ $2\ CuCl$ (Cuprous Chloride) $+$ $2\ HCl$ (Hydrochloric Acid)

(2) $2\ CuCl$ (Cuprous Chloride) $+$ $2HCl$ (Hydrochloric Acid) $+$ $1/2\ O_2$ (Oxygen) $\rightarrow$ $2CuCl_2$ (Cupric Chloride) $+$ H_2O (Water)

(1) $+$ (2) $2\ RSH$ (Mercaptan) $+$ $1/2\ O_2$ (Oxygen) $\rightarrow$ $RSSR$ (Disulfide) $+$ H_2O (Water)

As shown by the equations, one of the advantages of these processes is the theoretical absence of chemical consumption in the sweetening step. In practice it is necessary only to make up mechanical losses of the treating solution. Other operating charges are reduced, particularly labor costs, because of the simplicity of the method.

The elimination of gasoline loss, due to treating in a closed system and complete separation from the treating solution, contributes substantial savings. Probably the most important item is the improvement in quality of the copper-treated product. The product is also more responsive to gum inhibitors. The product is unusually stable when exposed to sunlight, with respect to both color and to corrosion test. Copper sweetening can form only the disulfides, since no additional inorganic sulfur is present in any form during the reaction. In the treatment the sulfur content is, of course, unchanged.

The Perco processes are applicable to any stock requiring sweetening and are especially suitable for those of a very sour nature since these are handled with only a slightly higher cost due to increased air requirements. This method of sweetening is applied in two ways; if the stock to be treated is sufficiently stable to color and gum when treated in presence of air, it is mixed with the necessary air and percolated through a bed of absorbent material impregnated with the copper reagent; other stocks are continuously treated with the copper solution in the absence of air and the copper regenerated in a separate tank. The oxidizing power of the solution is controlled to prevent side reactions with unstable hydrocarbon constituents. Stocks treated with the copper solution are given a final sodium sulfide wash to give color and gum stability, by removing minute traces of copper which may be present in the treated oil.

The simplified flow diagrams illustrate the two methods. In either process, if the gasoline to be sweetened contains hydrogen sulfide, this component is removed by a caustic-washing step prior to the sweetening operations. Likewise, if free sulphur is present in the stock, this compound is removed by a caustic-polysulfide treatment prior to the sweetening operation. If hydrogen sulfide and free sulphur are not present, these steps may be omitted.

In the Perco solution process, the gasoline is continuously treated by contacting with the treating solution in a centrifugal pump. The mixture of copper solution and sweetened gasoline passes to a separating tank, from which the used copper solution is continuously withdrawn to the regenerating tank where it is regenerated with a controlled amount of air to give the desired oxidation potential. The maximum efficiency of the air is obtained by diffusers, as illustrated. The regenerated copper solution overflows to a solution-storage tank and is then recycled to the treating step. The sweetened gasoline is continuously withdrawn from the top of the separating tank and is given the final stabilizing wash, settled, and passed to finished storage. In the solid process, the feed, free of hydrogen sulfide, is mixed with a predetermined quantity of air, which is dissolved in the gasoline, percolated through the reagent and passed to finished storage.

In the Perco solution process all equipment in contact with the copper solution is constructed of corrosion-resisting material, as the solution is corrosive to iron or steel. In the solid process ordinary materials are used throughout.

Due to compactness of equipment and relatively small size for a given throughput, the installation costs are low.

The following example of liquid treatment of a cracked distillate having a 170-400° F. boiling range and 53° API gravity is typical of results obtained in comparison to doctor treating.

	Sour	Doctor Treated	Perco Treated
Color, Saybolt	+2	−5	+2
Octane Number:			
Without Lead	62.9	61.9	62.9
+1 cc TEL	68.5	67.8	68.5
+2 cc TEL	71.7	71.0	71.7
Induction Period, Min.:			
Without Inhibitor	130	130	130
With 0.002 Percent Inhibitor	320	190	415
With 0.004 Percent Inhibitor	485	305	640
ASTM Gum	2.6	0.4	0.0

PERCO COPPER SWEETENING PROCESS
SOLUTION PROCESS
SOLID PROCESS
CAUSTIC CONTACTOR
CAUSTIC SETTLING TANK
CHARGE SOUR GASOLINE
COPPER CONTACTOR
AERATOR TANK
FLOW TANK
SEPARATING TANK
SULFIDE CONTACTOR
SULFIDE SETTLING TANK
SWEET GASOLINE
AIR
RV
GG
TREATING TOWER
AIR MIXER
AIR STRAINER
DUAL PROPORTIONING RECORDING FLOW CONTROL
COURTESY, PHILLIPS PETROLEUM COMPANY

Clay Treating Process

TEXACO DEVELOPMENT CORPORATION

THIS Process, named after its inventor, the late T. T. Gray, was first put into commercial operation in 1924 at Barnsdall, Oklahoma. The licensed Gray "clay treating" process is used in treating gasoline produced by various cracking, reforming, and polymerization processes. These Gray towers have treated the product of Cross, de Florez, Dubbs, Gyro, Holmes-Manley, Tank & Tube, Combination Units and others, from variegated charging stocks ranging from Pennsylvania gas oil to naphthenic base crudes.

The process is based on the selective polymerization of the unstable, unsaturated hydrocarbons resulting from pyrolysis. The gum forming and other objectionable properties of the gasoline are eliminated, while the desirable constituents pass through the process unchanged. The result is that treating losses are reduced to the vanishing point, with the octane number of the gasoline being processed unchanged. Most Gray installations are made in direct connection with the cracking unit whereby the rerunning step is eliminated and a treated gasoline is produced directly at the cracking plant. Vapors to be treated are passed through fuller's earth or similar adsorbent material, and the unstable unsaturated hydrocarbons are thereby polymerized into higher boiling compounds which are separated from the treated gasoline by virtue of their higher boiling point.

Treating costs are low. Operating costs are often only a fraction of a cent per barrel. No added labor is required and 3,000 to 30,000 barrels of gasoline have been treated per ton of clay before replacement or revivification is required.

The apparatus required is simple. Fuller's earth (usually 30-60 mesh) is charged to one or more of the treating towers (polymerizers) through which the vapors are passed. The greater part of the polymers formed separate from the vapors in the tower, and any remaining polymers may be removed by a dephlegmating or fractionating step before condensation of the treated vapors. Division of the fuller's earth into a series of beds gives better results than where the clay is used in one bed. However, except in cases where extremely low gum is required, a single treating tower is ordinarily sufficient. The flow diagram presented on the opposite page shows a typical arrangement where two polymerizers are employed. The towers are so manifolded that either can be used first, or, if desired, they can be used in parallel.

The polymerizers are usually operated at the condensing pressure of the distillation unit to which they are connected. It has been found that the higher the pressure, the greater the throughput per ton of clay, all other things being equal. Commercial units have been successfully operated at pressures as high as 375 to 400 pounds.

The quality of gasoline made during a run is quite uniform, since the decline in activity of the clay is gradual. Often, the activity of the treating material can be effectively prolonged by passing steam or a gasoline spray through the treating towers for a short time. Plants operate 90 days or longer between changes of clay.

Following are results obtained in a typical installation:

RESULTS ON REFINING A HOLMES-MANLEY CRACKED GASOLINE FROM KENTUCKY CRUDE

Cracking Plant	Holmes-Manley Redistilled Circulation
Charging	Topped Kentucky Somerset Crude (Gravity 28.4° API)
Clean Oil Transfer Temp.	965-980° F.
Two Gray Towers in series	174 lbs. pressure top 1st Tower
Type of Clay	Olmstead
Throughput when sampled	3627 barrels per ton
Tests on Gasoline:	
Gravity	61.7°F. API Color-Saybolt 30+
Distillation: IBP	84°F. Octane 78-77
10%	124°F.
20%	...
50%	230°F.
90%	357°F.
End-point	396°F.

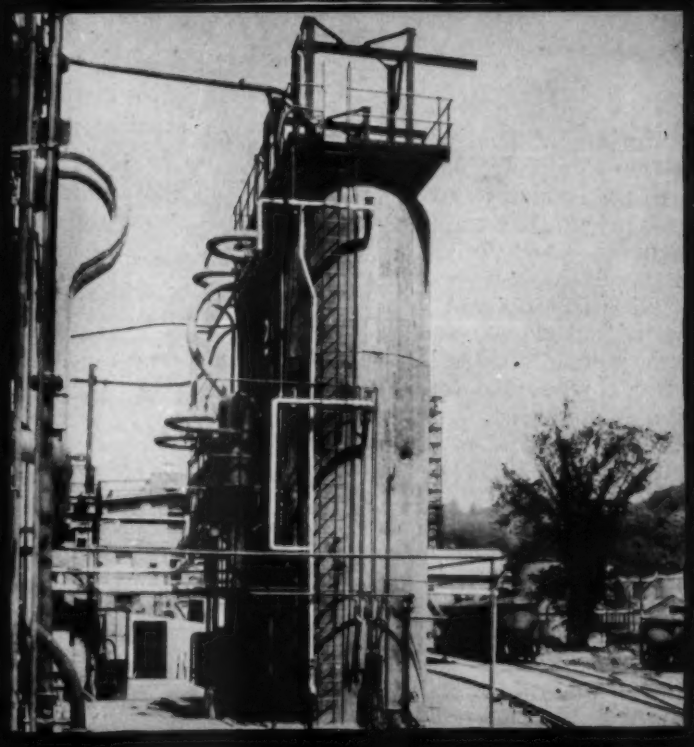
Gum Tests:	Doctor Sweetened
Copper Dish (mg./100 ml.)	3
Oxygen Bomb: Induction Period (hrs.)	2½

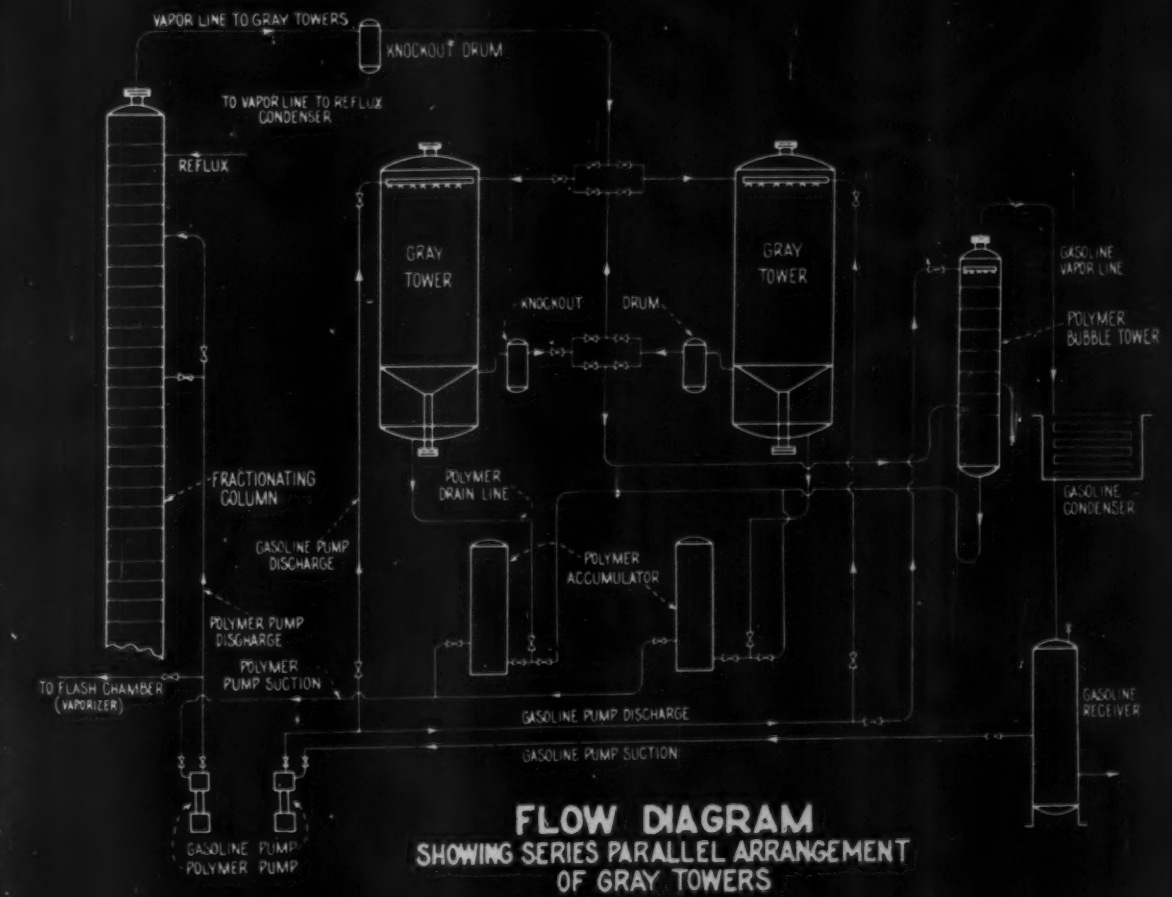
Storage Behavior of Gasoline (Doctor Sweetened Samples):

Months Stored	1	2	3
Copper Dish Gum (mg./100 ml.)	2	2	4
Color Saybolt	+30	+29	+29

Note the low copper dish gum test and good storage behavior.

The Gray Clay Treating Process is licensed by The Gray Processes Corporation, 26 Journal Square, Jersey City, New Jersey.





FLOW DIAGRAM
SHOWING SERIES PARALLEL ARRANGEMENT OF GRAY TOWERS

COURTESY, THE GRAY PROCESSES CORPORATION

Catalytic Desulfurization Process

PHILLIPS PETROLEUM COMPANY

THE Perco Catalytic Desulfurization Process developed by the Phillips Petroleum Company, Bartlesville, Oklahoma, is available to the industry through the Perco Division. The process provides an economical means of removing, in general, 40-60 percent of the sulfur content of cracked gasolines, 60-98 percent from straight-run gasolines, and 90-100 percent from natural gasolines. The types of sulfur compounds removed by the process include mercaptans, sulfides, and disulfides, while the cyclic compounds are not greatly affected. The variation in quantity of cyclic compounds present in the above mentioned types of gasoline accounts for the differences in reduction of total sulfur content.

It is well recognized now that sulfur compounds in motor fuels have a marked effect on the response of tetraethyl lead, even in fuels containing less than 0.05 percent sulfur. Basically, Perco Catalytic Desulfurization results not only in "desulfurization" but also in a "catalytic" effect resulting in octane and TEL response improvement not otherwise obtained merely by sulfur removal by conventional means. In many cases other important benefits are realized such as improvement in odor, improvement in product stability with respect to gums and color as well as reduction in corrosion to process and storage facilities.

Catalytic desulfurization is particularly applicable for improving the quality of naphthas to be used in the manufacture of high-octane aviation gasolines. Substantially complete desulfurization, as obtained in this process, is required for maximum response to tetraethyl lead.

As shown in the accompanying flow sheet, the feed is usually vaporized by heat exchange with the product leaving the catalyst tower, and then superheated to a temperature in the neighborhood of 750° F. The superheated vapors are passed through the catalyst bed, at fifty pounds pressure or less, where organic sulfur compounds are decomposed to hydrogen sulfide. The flow rate may range from 100 to 300 barrels of liquid feed stock per ton of catalyst per day. After cooling and condensing the product, the hydrogen sulfide is removed by whatever means is most applicable in the particular instance, such as fractionation or chemical treatment or a combination of both.

Operating cycles up to 30 days or more for a charge of catalyst are obtained on many gasolines. Catalyst cost is in the order of one cent per barrel of gasoline treated.

The treatment of cracked gasolines or blends containing cracked gasolines tends to deactivate the catalyst. Auxiliary equipment is provided for catalyst regeneration, which is accomplished by charging to the alternate chamber a stream of low-pressure air to burn off the carbonaceous matter. The air is admixed with a predetermined quantity of steam superheated to give a temperature of approximately 750° F. for the air-steam mixture entering the reactor on regeneration.

The rugged catalyst employed in the process can be regenerated an indefinite number of times without appreciable impairment of its activity. There is substantially no loss of hydrocarbons involved in the desulfurization treatment.

Several typical examples of the effects of catalytic desulfurization on lead response are shown in the accompanying tables for various types of gasoline.

CRACKED GASOLINE
56° API Gravity; 9-Lb. R.V.P.; 420° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	65.7	69.1
+1 cc TEL	70.0	75.8
+2 cc TEL	72.4	78.0
+3 cc TEL	74.5	79.8
Total Sulfur	0.140	0.076
Mercaptan Sulfur	0.062	0.003

STRAIGHT RUN NAPHTHA
50° API Gravity; 3-Lb. RVP; 440° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	45.4	49.6
+1 cc TEL	52.0	59.8
+2 cc TEL	56.6	66.8
+3 cc TEL	59.8	70.5
Total Sulfur	0.323	0.053
Mercaptan Sulfur	0.119	0.002

STRAIGHT RUN GASOLINE
58° API Gravity; 6-Lb. R.V.P.; 400° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	53.8	57.8
+1 cc TEL	59.2	68.9
+2 cc TEL	63.0	74.1
+3 cc TEL	65.9	77.4
Total Sulfur	0.309	0.017
Mercaptan Sulfur	0.148	0.000

STRAIGHT RUN GASOLINE
64° API Gravity; 8-Lb. R.V.P.; 375° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	58.5	60.8
+1 cc TEL	65.4	72.8
+2 cc TEL	69.4	78.5
+3 cc TEL	72.2	81.5
Total Sulfur	0.143	0.005
Mercaptan Sulfur	0.067	0.001

CRACKED-STRAIGHT RUN-POLY GASOLINE
66.2° API Gravity; 9.9 R.V.P. 355° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	58.6	63.9
+1 cc TEL	66.4	73.9
+3 cc TEL	73.1	79.0
Research Octane No.	61.4	67.2
+1 cc TEL	70.0	76.6
+3 cc TEL	77.9	83.7
Total Sulfur, Percent	0.088	0.024
Mercaptan Sulfur, Percent	0.023	Sweet

STRAIGHT RUN GASOLINE
68.3° API Gravity; 12.4 R.V.P. 346° F. E.P.

	Undesulfurized	Desulfurized
ASTM Octane No.	44.8	46.8
+1 cc TEL	53.9	59.6
+3 cc TEL	63.4	71.5
Total Sulfur, Percent	0.056	0.005
Mercaptan Sulfur, Percent	0.024	0.003

PERCO CATALYTIC DESULFURIZATION PROCESS
LEGEND
HYDROCARBON FLOW
REGENERATION GAS FLOW
STEAM
AIR
STEAM SUPERHEATER
TO VENT STACK
HYDROGEN SULFIDE
PRODUCT TO HYDROGEN SULFIDE REMOVAL
CATALYST CHAMBERS
EXCHANGERS
COOLER
ACCUMULATOR
PROCESS HEATER
FEED
COURTESY, PHILLIPS PETROLEUM COMPANY

Catalytic Desulfurization Process

TEXACO DEVELOPMENT CORPORATION

THE Gray catalytic desulfurization process, now commercially used for the elimination of both mercaptan and non-mercaptan sulfur compounds from lighter petroleum fractions, meets the need for a catalytic process utilizing a low-cost catalyst and enabling the employment of a highly selective procedure to minimize product losses through undesirable side effects.

To accomplish the desulfurization, gasoline vapors are passed through a bed of the catalyst at an elevated temperature, the temperature dependent on the type of charge stock and the degree of effect desired. The pressure of operation, while usually of a low range, is sufficient to secure the desired flow of vapors through the process and auxiliary equipment.

The catalyst used is of the solid adsorbent type, such as fuller's earth and the like, which is low in cost and readily available from many widely spaced deposits in the United States, and therefore conveniently accessible to all refining centers in this country. It has been found that a wide selection of effective clays can be used and several plants operating the Gray catalytic desulfurization process are using different fuller's earth, choice of clay being made after consideration of operating and economic factors. A flow diagram is shown.

In the application of the process to straight-run gasoline (referring to the flow diagram), the desired fraction is heated to a temperature in the order of 700° F. in a furnace of simple design. The heated vapors pass through a bed of solid adsorbent catalyst at a flow rate dependent upon the character of the feed and nature of the sulfur compounds, wherein the conversion of mercaptan and non-mercaptan sulfur compounds takes place. The vapors leaving the catalyst chamber pass in heat exchange with the incoming gasoline and go then to a condenser and separator. A large part of the sulfur present in the original mercaptan and non-mercaptan sulfur compounds is converted to hydrogen sulfide, which is partially removed in the separator and the remainder removed by means of either a caustic wash or simple stabilization. One of the important economic factors realized as a result of application of this Gray process to straight-run gasoline is that, other than removal of hydrogen sulfide, no further treatment of the gasoline is required.

The desulfurization of 100% cracked gasoline ordinarily requires that the gasoline be pre-treated by the conventional Gray treating process prior to desulfurization in order that the most economical catalyst life be realized. Catalyst yields here range upward of 1,200 bbls./ton. The use of the conventional Gray process prior to desulfurization can be additionally justified on the basis that it effects refining of the gasoline by the removal of diolefins and thereby obviates the need for using inhibitors in the final product.

Untreated cracked gasoline is being desulfurized in a commercial installation with substantial lead savings in the form of a 50-50 blend with straight-run gasoline. In this case the straight-run gasoline, as well as the cracked, has poor lead susceptibility, and both cracked and straight-run components are together improved by desulfurization treatment.

Catalyst yields range from 800 to 1000 barrels per ton for this operation and the low cost of the fuller's earth type catalyst used in the Gray desulfurization process makes it possible to realize substantial net savings over the cost of operation at these catalyst yields.

Licensed by The Gray Processes Corporation, 26 Journal Square, Jersey City, New Jersey.

Desulfurization of Various Gasolines

	Desulf. of S. R. Gasoline from New Mexico Crude**		Desulf. of S. R. Gasoline from Michigan Crude**		Desulf. of Gasoline from Illinois Crude**		Desulf. of 50-50 Blend of Untreated Cracked Gasoline and Straight Run Gasoline from Michigan Crude***	
	H_2S Free Charge	H_2S Free Product	H_2S Free Charge	H_2S Free Product	H_2S Free Charge	H_2S Free Product	H_2S Free Charge	H_2S Free Product
Gravity, A.P.I.	55.5	55.6	67.8	67.6	.....	.....	64.8	64.2
Distillation:								
IBP	115	114	98	98	132	131	98	100
10 percent	169	173	135	134	188	187	136	140
50 percent	266	261	218	216	263	261	228	230
90 percent	387	397	290	286	349	348	320	330
EP	436	442	338	342	396	395	360	384
ASTM Gum	1	2	.....	.....	.....	.....	1.6	2.0
	Percent	Percent	Percent	Percent	Percent	Percent	Percent	Percent
Total Sulfur	0.138	0.0280	0.053	0.005	0.038	0.006	0.105	0.042
Mercaptan Sulfur	0.057	0.0017	0.018	0.0004	0.007	0	0.040	0.000
Non-Mercaptan Sulfur (by diff.)	0.081	0.0263	0.035	0.0046	0.031	0.006	0.065	0.042
Doctor Test	Positive	Positive*	Positive	Negative	.....	.....	Positive	Negative
ASTM Octane No.:								
Clear	54.4	56.6	48.8	49.4	47.8	50.3	58.8	60.4
Plus 1cc TEL	61.3	67.2	59.2	62.6	58.1	62.4	65.1	67.0
Plus 2cc TEL	66.3	73.0	63.9	69.0	63.7	70.0	67.8	71.0
Plus 3cc TEL	69.4	77.2	67.0	73.1	68.0	74.0	70.3	74.3
1939 Research Octane No.								
Clear	56.5	59.6	.....	.....	.....	.....	.....	.....
Plus 1cc TEL	63.4	68.7	.....	.....	.....	.....	.....	.....
Plus 2cc TEL	68.8	73.4	.....	.....	.....	.....	.....	.....
Plus 3cc TEL	71.5	76.9	.....	.....	.....	.....	.....	.....
HR&G Lead Susceptibility	1.19	1.8	1.31	1.90	1.57	2.10	0.9	1.15
Clear Octane Increase:								
ASTM	.....	2.2	.....	0.6	.....	2.5	.....	1.6
1939 Res	.....	3.1	.....	.....	.....	.....	.....	.....
TEL Required for 63 O. N	.....	.....	2.35cc	1.35cc	.....	.....	.....	.....
TEL Required for 68 O. N	.....	.....	.....	.....	3.0	1.7	.....	.....
TEL Required for 69 O. N. ASTM	2.80cc	1.25cc	.....	.....	.....	.....	.....	.....
TEL Required for 70 O. N	.....	.....	.....	.....	.....	.....	3.0cc	1.6cc
TEL Required for 70 O. N. (Extr.)	.....	.....	4.20cc	2.20cc	.....	.....	.....	.....
TEL Required for 71 O. N. 1939 Res	2.70cc	1.40	.....	.....	.....	.....	.....	.....
TEL Required for 72 O. N. (Extr.)	.....	.....	.....	.....	.....	.....	4.0cc	2.2cc
TEL Saving at 63 O. N	.....	.....	.....	1.0	.....	.....	.....	.....
TEL Saving at 68 O. N	.....	.....	.....	.....	.....	1.3	.....	.....
TEL Saving at 69 O. N. ASTM	.....	1.55cc	.....	.....	.....	.....	.....	.....
TEL Saving at 70 O. N	.....	.....	.....	.....	.....	.....	.....	1.4cc
TEL Saving at 70 O. N. (Extr.)	.....	.....	.....	2.0	.....	.....	.....	.....
TEL Saving at 71 O. N. 1939 Res	.....	1.30cc	.....	.....	.....	.....	.....	.....
TEL Saving at 72 O. N. (Extr.)	.....	.....	.....	.....	.....	.....	.....	1.8cc

* Negative after Caustic Wash. ** Catalyst Life Approximately 5,000 barrels per ton. *** Catalyst Life 800 barrels per ton.

CLAY DESULFURIZATION UNIT
CLAY TOWER
EXCHANGER
CONDENSER
COOLER
ACCUMULATOR DRUM
H2S GAS TO HEATER
GASOLINE TO H2S WASHER
TRANSFER PUMP
CHARGE PUMP
HEATER
CHARGE IN
COURTESY GRAY PROCESSES CORPORATION

Don't penalize operating efficiency by deferring cleaning TOO LONG!

When carbonized oil, sludge, wax, lime-scale, rust, slime and other deposits have completely CLOGGED your shell and tube type equipment you are immediately confronted with a tough, time-consuming, expensive cleaning job to restore heat transfer.

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But there is an easy way you can avoid this situation. DON'T POSTPONE OR DEFER CLEANING TOO LONG! Instead, clean your heat exchange equipment on a regular periodic basis. Doing this saves time, cuts cleaning cost. Three effective Oakite cleaning methods are available to you for this important and continuing maintenance work: (1) Circulating method; (2) Tank method; (3) Steam cleaning method. Take advantage of specialized Oakite materials and long years of Oakite experience in helping other refineries set up this maintenance work on a planned schedule of cleaning because heat exchangers can then be cleaned FASTER, deposits removed EASIER and down-time considerably SHORTENED.

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This 28-page Digest, based on the successful use of Oakite materials and methods in the producing, refining, and marketing divisions, describes 88 different maintenance cleaning jobs. Write for your free copy today.

Gas Purification, Dehydration Processes

DURING the last decade the tremendous increase in the consumption of natural gas as industrial and domestic fuel has resulted in the construction of many long distance gas transmission pipelines of large diameter. This demand for natural gas has drawn heavily on the readily available sweet gas production, and has forced the utilization of gases once rejected as unsuitable because of excessive hydrogen sulfide contamination. The same period has witnessed, also, the more general utilization of high sulfur crude oils for refinery feeds, and the more frequent presence of serious amounts of hydrogen sulfide in natural gasoline plant feeds.

The large reserves of sour natural gases, the needs of process plant operators, plus excellent markets for all petroleum products, has focussed attention on processes for hydrogen sulfide removal. In these processes the resulting purified gas may have a hydrogen sulfide content at almost any specified level to suit the requirements of the specific application. Thus great volumes of gas formerly flared because of toxic and corrosive properties are now used, free of hydrogen sulfide, to alleviate fuel gas shortages in many parts of the nation and real savings are made in process plant maintenance costs.

Dehydration of gases moved by long distance pipe lines is a necessity if efficient transmission is to be realized. The provision of equipment for the dehydration of the gas to prevent hydrate formation, water plugging, and corrosion in pipe lines is now accepted as standard design and is regarded almost as important as compressor stations.

The separate processes for the dehydration and the desulfurization of sour gases through the use of liquid absorbents were developed hand-in-hand. In recent processes these steps have been merged for simultaneous operation. The absorbents used for desulfurization and removal of carbon dioxide from sour gases have been combined with dehydration absorbents to effect important savings in equipment investment and operating costs.

Early dehydration plants mainly used a system by which solid desiccant beds functioned to adsorb moisture from the gases treated. Such processes are still extensively used and have certain important advantages. Regeneration of the adsorbent is accomplished in place by means of a stream of hot gases or air. Keen competition exists in the gas dehydration field between adsorption and absorption type processes in cases where gas purification is not a factor. Both systems require regeneration of the dehydrating material and have their individual merits and problems.

Investment and operating costs of dehydration and purification processes are not subject to easy generalization, due to variations in plant size, hydrocarbon composition of the gas to be treated, amounts of moisture and/or hydrogen sulfide content, pressure of the treated gas, utility costs, and factors of plant location with relation to other process facilities. Costs are usually modest when weighed against possible gains. Many purification plants are used as feed preparation units for refinery conversion processes, where the deleterious effects of hydrogen sulfide contaminants on a catalyst may be the major factor governing the installation. Similarly, many gas dehydration units are used ahead of low temperature processing operations where the removal of substantially all moisture content is essential to the successful operation of the subsequent process step. Whatever the reasons for the installation of specific plants, the development of such processes has had the very important result that nearly all available gas supplies, previously rejected for hydrogen sulfide content, now can be used to good advantage.

Adsorptive Gas Dehydration Process

HOUSTON OIL FIELD MATERIAL COMPANY

ONE of the contributing factors in the increasing popularity of the adsorptive type drier is the simplicity of design and method of operation. The gas to be dried is passed through a vessel containing a granular desiccant until the desiccant is saturated with moisture, then heating the desiccant to drive off the moisture, thus restoring it to its original condition.

As shown in the diagram, wet gas is first passed through a separator at a controlled temperature and pressure to remove any liquid—oil, distillate or water—that may be present. This reduces the liquid load on the desiccant and also precludes the possibility of fouling it.

The dehydration unit, being of the pressure differential type, requires the use of pressure controlling equipment between the separator and the inlet manifold of the contactors. Diaphragm operated motor valves or adjustable chokes are used at this point. The volume of gas flowing through the unit is controlled here in conjunction with the pressure.

The gas from the separator leaves the control point at a predetermined pressure and enters the contactor through the inlet manifold. The gas flows through the desiccant bed from the top of the contactor downward. The desiccant is supported by two or more trays with a free space between the beds for redistribution to minimize the possibility of channeling.

An activated bauxite is used, which is a porous, granular, industrial and laboratory drying agent for liquids and gases. It adsorbs water instantly, but is insoluble and will not swell, disintegrate in water, or appear wet at the end of the adsorptive part of a cycle. The granules are hard, neutral, stable, noncorrosive, and can be handled freely by hand without harm. Repeated regeneration does not change its composition. After contact with this desiccant, a gas may encounter extremely low temperatures without the formation of hydrates or condensation of water. With good regeneration and favorable operating conditions, even greater drying efficiency is obtainable. It has the ability to adsorb from twenty to twenty-five percent of its weight of water from gases before regeneration is necessary.

From the desiccant bed in the contactor, the gas, which is now dry, flows through the outlet manifold to the heat exchanger. The gas flows through the shell side of the exchanger for the purpose of cooling the regeneration gas which is passing through the tubes. From the exchanger the dry gas flows through the measuring station into the transmission line.

Upstream from the pressure control point a portion of the gas is diverted from the main stream for regeneration purposes. The pressure and quantity of this gas is controlled by means of a diaphragm operated valve or an adjustable choke. The gas flows from this control point to the heating unit where it is heated to a temperature of 400° F. This heated gas flows through the saturated desiccant bed from the bottom upward, revaporizing the liquid adsorbed and carrying it out of the contactor. Since the gas being dried flows through the desiccant bed from the top downward, the liquid content is higher in the top at the first point of contact. Therefore the flow of the regeneration gas is reversed, flowing through the bed from the bottom upward, carrying the revaporized liquid off the top. This method reduces the time necessary for regeneration, because the liquid does not have to be driven through the entire desiccant bed.

When the regeneration gas leaves the contactor being dried, it passes through the tube side of the heat exchanger to be cooled by the cold gas in the shell. The cooled gas flows through a second separator to remove most of the liquid taken out of the contactor being dried. The gas from this separator is introduced into the main gas stream on the downstream side of the main pressure control point with sufficient pressure to force it into the stream, so that it can flow through the drying contactor with the wet gas and be fully dehydrated. This pressure is controlled at the control point in the regeneration line.

Temperature controllers are located in the heated gas line at the outlet of the heater, and in the regeneration gas line at the outlet of the contactor, to assure proper temperature control. A flow controller is located in the regeneration line.

BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
ADSORPTIVE GAS DEHYDRATION PROCESS
COURTESY, HOUSTON OIL FIELD MATERIAL COMPANY
CONTACTOR
CONTACTOR
SEPARATOR
SEPARATOR
HEAT EXCHANGER
HEAT EXCHANGER
STEAM
SATURATED GAS
LIQUID REMOVED FROM GAS
DRY GAS
2000000
1100
90
1050
65
1115
70-375
1140
400
1025
70
1105
100
1150
70
1200
75
20000000
1000
90
20000000

Diethylene Glycol Gas Dehydration Process

NATURAL gas normally contains a certain amount of water vapor. Unless the water content is reduced so that no condensation occurs under the conditions of temperature and pressure existing in natural gas transmission lines, operating difficulties are likely to develop. Moisture causes internal corrosion and rust accumulations. Water condensing at the low points in the transmission line must be drained off from time to time to prevent stoppage and freezing. In addition, at high pressures and in the presence of liquid water, solids known as natural gas hydrates may form, which resemble snow and build up in the line, seriously interfering with its capacity. At high pressures these hydrates may have melting points as high as 50-60° F. It is therefore apparent that atmospheric temperature conditions do not have to be particularly severe before difficulties may be encountered.

For the above reasons it has been necessary to provide for dehydrating natural gas to reduce the dew point before sending the gas through long high pressure transmission lines which are subjected to low atmospheric temperatures. Dehydration to a dew point below the minimum temperature to which the line will be subjected in the territory through which it passes has the following advantages:

(1). Maintenance at all times of full line capacity which would otherwise be reduced due to the formation of hydrates and water condensate. (2). Elimination of the necessity of blowing the drips to remove condensate. (3). Elimination of regulator freeze-ups. (4). Reduction of internal corrosion of the pipe line.

Natural gas dehydration is generally accomplished by contacting the gas with an hygroscopic liquid. Refrigerated calcium chloride water solutions have been used for this purpose. The cycle of operation when using diethylene glycol is basically the same as for brine. Dew point depressions up to 60° F. can be secured with concentrated diethylene glycol without refrigeration, which is necessary when correspondingly low depressions are to be secured with brine solutions. Refrigeration equipment materially increases plant investment. Diethylene glycol is not corrosive, and has a low freezing point so there is no possibility of solidification of the solution in the system.

Properties of Diethylene Glycol

Formula	$HOCH_2CH_2OCH_2CH_2OH$
Molecular Weight	106.08
Boiling Point	244.5°C. (472.1°F.)
Freezing Point	—6.5°C.
Specific Gravity	1.1185 20/20°C.
Weight Per Gallon	9.308 lb. 20°C.
Viscosity	0.38 poises 20°C.
Vapor Pressure	0.015 mm. Hg 20°C.
Specific Heat	0.5509 20°C.
Latent Heat of Evaporation	150 cal./g. at B. P.
Coefficient of Expansion (per °C.)	0.000635 at 20°C. 0.000650 at 55°C.

Operation of the plant, as shown in the flow diagram on the opposite page, is as follows: Cooled diethylene glycol containing a small amount (generally less than 10 percent) of water depending on the final dew point desired is picked up from the solution storage tank (1) by a pump (2) capable of pumping against the pressure at which the gas is to be treated. The solution flows through a flow controller (3) which controls the speed of the pump and into the top of the tray type contactor (4). Gas enters the bottom of the contactor through a mist extractor to remove mineral seal oil, etc. The contactor is equipped with bubble cap trays, the solution descending counter-current to the ascending natural gas. The gas leaves the contactor through mist extractor (6) and outlet connection (7) into the transmission line. The solution which has absorbed water from the gas is collected on the chimney plate (8). The liquid level on the chimney plate is controlled by the flow controller (9) and flows into a vent tank (10) provided with a back pressure regulator (11) which maintains sufficient pressure in the vent tank to force the liquid through the heat exchangers (12) and the reflux condenser (12-A) into the top of the still (13). Natural gas released in (10) is vented through the vent gas outlet.

Glycol solution passes through the reflux condenser (12-A) into the still through valves 14, 15 or 16 as desired, generally through one of the lower openings. The solution flows down over bubble cap trays in the still column and into the lower portion of the re-boiler (17) where it is heated by closed steam coils and is returned to the kettle (18) in the lower part of the still. Water remaining in the heated solution is vaporized in the kettle and passes through the bubble cap trays of the still column. This serves to heat the descending solution and separates the diethylene glycol from the water vapor, thus eliminating loss of diethylene glycol in the still. Any natural gas or uncondensed water vapor in the reflux condenser (12-A) are vented through the back pressure valve (20). Excess condensate in the reflux condenser is drained through the condensate line to the sewer. The solution in the kettle (18) having been substantially denuded of water passes through the heat exchangers (12) in counter-current relation to the weak solution from the vent tank, and thence through a cooler (19) to storage tank (1). Replacement solution is added to the system when necessary, at this point. The cycle of operation described is continuous.

In commercial plants the total over-all loss of diethylene glycol averages .05 gallons per million cubic feet measured at atmospheric pressure.

Diethylene glycol is manufactured by the Carbide and Carbon Chemicals Corporation, 30 East 42nd Street, New York, N. Y.

DIETHYLENE GLYCOL GAS DEHYDRATION PROCESS
BACK PRESSURE REGULATOR
WATER CONDENSATE
STILL
266
25
STEAM
295
HEAT EXCHANGER
SOLUTION STORAGE TANK
VENT
VENT TANK
MIST EXTRACTOR
MIST EXTRACTOR
SOLUTION PUMP
DEHYDRATED GAS (28°F. DEW POINT)
CONTACTOR
MIST EXTRACTOR
WET GAS IN (65°-75°F. DEW POINT)
20,000,000
725
5
B BARRELS PER DAY
G GALLONS PER MINUTE
P POUNDS PER HOUR
M MOLS PER HOUR
S STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

Girbotol Process

THE GIRDLER CORPORATION

THE Girbotol Process (patented) for the separation of acidic constituents from gaseous and liquid mixtures, is widely used for purification of gases and liquids in the petroleum industry. The process is employed for removal and recovery of hydrogen sulfide from natural and refinery gases, for simultaneous desulfurization and dehydration of natural gas, for purification of liquid hydrocarbons, and for separation of carbon dioxide from hydrogen used for hydrogenation of oil and synthesis of iso-octane.

Operation of the Girbotol Process is based upon the discovery that aliphatic amines have a strong affinity for acid gases such as hydrogen sulphide and carbon dioxide at or near atmospheric temperature, but at higher temperatures the affinity drops off rapidly and the gas is expelled. The process operates in a continuous cycle, absorbing acid gas from a gas stream at atmospheric temperature and expelling it at another point by raising the temperature of the absorbent. Because the amines are strong bases, freely soluble in water, large volumes of acid gas may be removed per volume of solution circulated. The efficiency of the process is therefore high. Hydrogen sulphide is recovered in concentrated form.

The cycle and equipment employed in the process are almost identical with those of a gasoline absorption plant (see flow chart). Gas to be treated flows up through a bubble tower or absorber. Amine solution, at atmospheric temperature, enters the absorber and flows down the tower from tray to tray picking up the acid gas. Purified gas leaves the absorber at the top.

Amine solution, saturated with acid gas, passes through heat exchangers, where its temperature is raised to 190-200° F., and into the upper part of the reactivator, which also consists of a bubble tower, containing in its base a tubular reboiler section and at the top a reflux condenser. The amine solution, flowing down the tower, is heated to 215-220° F. by steam rising from the boiling solution in the reboiler section, and the acid gas is expelled and flushed out by ascending vapors. The acid gas leaves the reactivator saturated with water vapor at 190-200° F. and flows through the condenser where it is cooled by water and the steam condensed. The condensate is returned to the reactivator. Hot, lean solution, stripped of acid gas flows from the base of the reactivator to a pump, thence through the exchangers in counter current relation to the cold, rich solution. From the exchangers the solution passes through water coolers, and finally back to the top of the absorber. The plant is entirely automatic in operation, requiring only casual supervision.

Desulphurization — Separation of hydrogen sulphide from hydrocarbon gases is accomplished as described above. The gases may be treated at high or low pressure with equal success, but gas under pressure is purified more economically because smaller towers are required and larger volumes of hydrogen sulphide are removed per gallon of solution circulated. The process is well adapted for purifying gas containing a high percentage of hydrogen sulphide. The rate of reaction in the absorption stage is so rapid that complete stripping may be obtained in a single tower. Because an aqueous absorbing solution is employed, practically no hydrocarbon constituents are removed from the gas if the physical conditions of temperature and pressure are controlled. This makes it possible to apply the process for the purification of wet gas before the extraction of gasoline and results in lower subsequent treating costs to purify the gasoline.

The Girbotol Process is useful for the purification of gaseous feed to a polymerization plant. Unless the poly feed is substantially free of hydrogen sulphide, mercaptans and other sulfur compounds formed during polymerization must be removed from the poly product. Such after-treatment is expensive and damaging to the octane number and the blending value of the poly product. The poly feed may be purified with the Girbotol Process at any temperatures up to 180° F.

Dehydration and Desulphurization—In several Girbotol plants purifying natural gas for pipe line distribution, simultaneous dehydration of the gas is accomplished by employing a dehydrating liquid such as diethylene glycol in solution with the Girbotol absorbent. H_2S and water are both scrubbed from the gas, and expelled together from the Girbotol reactivator, the H_2S assisting in carrying the excess water from the tower.

It is thus possible to sweeten sour natural gas and dry it in a single, simple plant. Such gas is being distributed through pipe lines to domestic and industrial customers, without troubles from pipe line freeze-ups.

Desulphurization of Liquid Hydrocarbons—The removal of hydrogen sulphide from liquid hydrocarbons, such as propane, butane, and gasoline, is being carried out successfully in a number of Girbotol plants. Packed towers are used for contacting the liquid streams, while the amine solutions are reactivated in the usual manner.

Carbon Dioxide Removal from Hydrogen—The Hygirtol Process for the production of high-purity hydrogen from hydrocarbons and steam utilizes the Girbotol process for the removal of carbon dioxide from hydrogen. The Girbotol process is the only cyclic process in commercial use for effecting complete removal of carbon dioxide from other gases at or near atmospheric pressure.

This information is presented through cooperation of The Girdler Corporation, Louisville, Kentucky, owner and licensor of the Girbotol Process.

GIRBOTOL PROCESS
ACID GAS TO DISPOSAL OR RECOVERY
STEAM
REFLUX DRUM
RICH SOLUTION
LEAN SOLUTION
PURIFIED PRODUCT
RAW FEED
1 MM
200
7
120
5
220
10
350
100
20
180
160
300
100
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
COURTESY THE GIRDLER CORPORATION

Glycol-amine Gas-Treating Process

THE FLUOR CORPORATION, LTD.

THE glycol-amine process for the purification and dehydration of a gas represents a continuous operation of two established processes. Aqueous solutions of the ethanolamines are capable of absorbing CO_2 and H_2S at atmospheric temperatures. When the solutions containing the acidic constituents are heated to their boiling points, these components are expelled from the solution. Similar use of diethylene glycol in the removal of water vapor from a gas stream led to a study for the combination of these two processes for the purification and dehydration in one unit.

The combination of these two principles as developed by The Fluor Corporation, Ltd., provides a process for simultaneous acid gas removal and dehydration, wherein the ethanolamine and diethylene glycol are included in the same solution. However, the combination feature, whereby both purification and dehydration may be accomplished in one stage is not the only advantage of the process. A glycol-amine solution may be regenerated more completely than an aqueous amine solution, thus permitting the attainment of higher purities in the gas being treated. At the same time, a glycol-amine solution with the same water content as a diethylene glycol solution will dehydrate the gas to a slightly greater degree than the latter.

The flow diagram on the opposite page represents a typical application in the treatment of a sour-gas stream. Sour gas is introduced into the bottom of the absorbing column, the contactor. As it passes upward through the bubble trays it is thoroughly contacted with the glycol-amine solution, which enters at the top of the vessel. The outlet gas, free of the CO_2 and H_2S and dehydrated, leaves the top of this column. The rich solution containing the absorbed acid gases leaves the bottom of the contactor and passes through a bank of conventional heat exchangers, wherein the temperature of the solution is raised. The rich solution then enters the regenerating column, the still. As it flows down over the bubble trays in this column it is contacted with the rising stripping vapors, thus stripping the solution of the acidic constituents. The solution from the bottom tray of the still enters a reboiler, wherein it is boiled to provide the stripping vapor for the column. The overhead stream of acid gas and water vapor from the still passes through a condenser, and the resultant mixture is separated in an accumulator, the acid gases passing overhead from this vessel and the condensate being returned to the still as reflux. The regenerated lean glycol-amine solution leaves the bottom of the still, passes through the exchangers, and is pumped back through coolers to the top of the contactor, thus completing the cycle.

When using an aqueous monoethanolamine solution it is possible to operate the still at higher than atmospheric pressures and gain somewhat more complete regeneration insofar as CO_2 is concerned, although increased pressure usually is deterimental in a stripping operation. In this particular case the dissociation of the monoethanolamine-CO_2 complex increases sufficiently with the increased temperature caused by the additional pressure to overcome the adverse effects of the increased pressure. Such is not the case with H_2S and the stripping of this constituent is less complete at an elevated pressure.

The addition of glycol to the solution increases the boiling point without necessitating an increase in pressure in the regenerating column. Thus, in the case of a glycol-amine solution, the adverse features of increased pressure are not encountered, and both CO_2 and H_2S may be stripped completely from the solution.

The maximum purity obtainable on any gas is a function of the pressure and temperature at the top of the acid-gas-absorbing column and the amount of acid gas remaining in the stripped solution. It is obvious that higher purities are attainable by the use of the glycol-amine solution than by the use of an aqueous monoethanolamine solution, since the former may be more completely stripped. This is particularly important when considering low H_2S specifications for treated gas. A difference of but a few grains of H_2S in the regenerated solution may mean the difference between meeting and not meeting the specified purity. The fact that glycol-amine solutions may be so easily stripped is one of the major advantages of the process. A corollary to the ease of stripping is the advantage that the utility requirements for a glycol-amine plane are in general less than those for an aqueous plant for comparable service.

As a result of the ready regeneration of the glycol-amine solution, this process is frequently used where the treated gas purity requirements are stringent, whether or not simultaneous dehydration is desired. Adjustment of the concentrations of the glycol and water permits setting the regeneration temperature to obtain the required degree of acid-gas stripping to insure the specified purity.

Typical solution compositions employed in the glycol-amine process may be comprised of from 10 to 30 percent monoethanolamine, 45 to 85 percent diethylene glycol, and 5 to 25 percent water. The ability of adjusting components of the glycol-amine-water solution to attain required results permits flexible operation. A wide range of specifications for treated gas purity and water content may be met by using this process. Sour natural gas may be treated readily so that the residue gas will contain less than 0.25 grains of H_2S per 100 cubic feet, which is less than 0.0004 mol-percent of H_2S in the gas.

As a typical example of the extremely high purity gas obtainable with the glycol-amine process under suitable operation there may be quoted data on a large operating installation in which the H_2S content of the gas has been reduced from 450 grs. per 100 SCF to less than 0.01 gr. per 100 SCF, or less than 0.16 mol parts per million.

This process is licensed by The Fluor Corporation, Ltd., Los Angeles.

GLYCOL-AMINE GAS TREATING PROCESS
COURTESY: THE FLUOR CORPORATION, LTD.
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
SOUR GAS INLET
50 000 000
H_2S - 2%
CO_2 - 3%
H_2O - SAT.
FLUOR SCRUBBER ELEMENTS
200
80
CONTACTOR
PURIFIED GAS
47 500 000
H_2S (@ 0.0004% OR @ 25 GR/100 SCF
CO_2 NIL
H_2O (@ 80 LBS/MMSCF
EQUIV. TO SAT. AT 35°F
80
430
GLYCOL-AMINE SOLUTION
5% H_2O
10-30% MEA
BAL. DEG
17 MM BTU/HR
COOLING WATER
20 MM BTU/HR
STILL
185
8
320
8
23 MM BTU/HR
STEAM (150 LBS)
27 000
5 MM BTU/HR
COOLING WATER
ACID GAS TO DISPOSAL
H_2S, CO_2, H_2O

Phosphate Desulphurization Process

SHELL DEVELOPMENT COMPANY

THE PHOSPHATE Process is a continuous, regenerative liquid purification system for the removal and recovery of hydrogen sulfide from natural and refinery gases and liquid hydrocarbon fractions. Treatment comprises two steps: absorption and regeneration. The principal absorption reaction may be expressed:

$$K_3PO_4 + H_2S \rightarrow K_2HPO_4 + KHS$$

Regeneration of the solution reverses this reaction.

Treatment of both the gas and liquid phases is identical with the exception of (1) extractor design, and (2) use of split solution flow for extraction of H_2S from the gas phase in difficult cases.

Considering Figure 1, the flow scheme depicting vapor phase extraction, the extraction step is carried out in a bubble tray column by means of split solution flow. The raw gas is contacted countercurrently with the lean solution in the lower section of the column to remove the bulk of the H_2S, and before its effluence as desulfurized gas it is further contacted with very lean solution to achieve the final degree of H_2S removal required. The combined solution streams issue from the bottom of the column, and after heat exchange with hot lean solution, this fat solution enters on the top tray of the regeneration column. In the regeneration column, the fat solution is stripped of H_2S by countercurrent flow of steam rising from the base of the column. Steam from an external source may be used, as in open steam stripping, or the steam may be generated by conventional reboiling of the solution, as shown. The lean solution, approximating 75 to 85 percent of the total solution flow, is removed from a sump located somewhat below the center of the column. A reconcentrator serves to maintain the proper concentration of this solution by boiling off the steam condensed in increasing the sensible heat of the solution. The remainder of the solution, the very lean solution, is further stripped to a lower H_2S level before leaving the column at the reboiler. For those cases in which a low residual H_2S content in the treated gas is required, this split solution flow system often results in steam savings of 25-40 percent over single flow treatment. Due to the relatively low flow rate of very lean solution, bubble tray design is most advantageously applied in design of split solution flow systems.

Steam and H_2S pass from the top of the regeneration column to a condenser, where steam is condensed, and the H_2S gas is expelled from the system. A simple condensing system consists of a partial condenser, located at the top of the column, which effects the desired degree of cooling of the H_2S yet supplies a hot condensate essentially free of H_2S which is returned directly to the reboiler from a condensate well, as shown in the flow diagram. In large installations a conventional condenser—accumulator system may be utilized, giving a cooled condensate which is returned to an additional stripping section at the top of the regenerator to remove H_2S. The lean solution leaving the regeneration column, after heat exchange with the fat solution, is cooled prior to being pumped to the extraction column. The smaller very lean solution stream normally does not exchange heat but is cooled directly and pumped to the extraction column.

Considering treatment of a liquid phase hydrocarbon, shown in Figure 2, the extraction step is achieved with countercurrent, single solution flow. Packed columns are usually specified but under certain conditions it is advantageous to use perforated plate columns. After extraction, the liquid hydrocarbon flows through a solution coalescer, which serves to safeguard against solution loss in the event of carryover or entrainment in the extraction column overhead. The fat solution flows from the bottom of the column to the fat solution settler, which serves to minimize hydrocarbon loss by entrainment in the solution. Subsequent to settling, the fat solution exchanges heat with the lean solution and flows to the regeneration column.

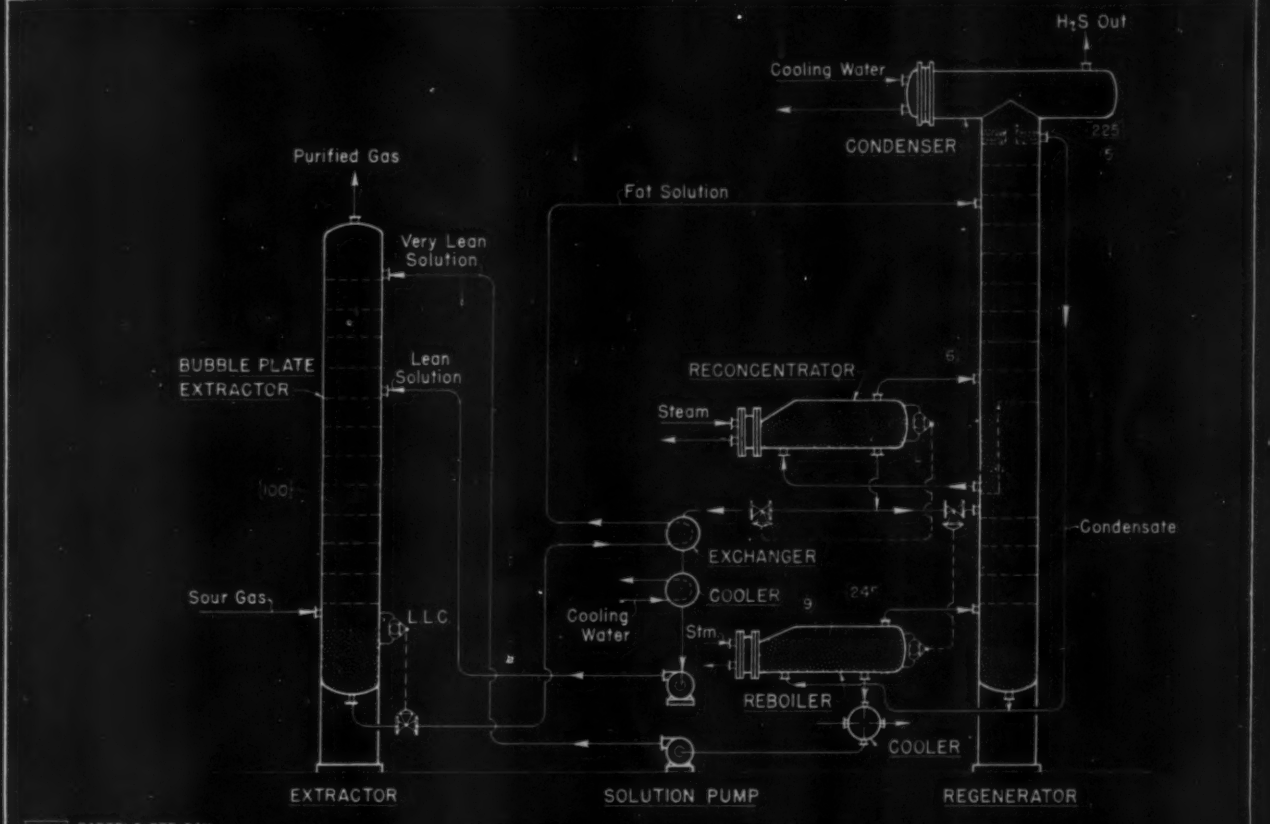
The regeneration column consists of a steam stripper. Fat solution enters on the top tray and is stripped by a countercurrent flow of steam issuing from the column bottom. Either open steam or reboiling may be used as a steam source for stripping. An additional method of condensation, i.e., direct condensation utilizing raw water, has been shown for the liquid phase extraction. However, the other methods are applicable. The resulting lean solution, after heat exchange with the fat solution, is cooled and pumped to the extraction column.

The hydrogen sulfide recovered from this process is essentially free of hydrocarbons and is an ideal raw material for the manufacture of sulphuric acid or elemental sulphur.

The inexpensive solution used ordinarily is prepared at the plant from commercial grades of potassium hydroxide and phosphoric acid. The optimum concentration of K_3PO_4 depends upon the amount and nature of impurities in the hydrocarbon processed. It does not contaminate nor polymerize the hydrocarbon. Also, since K_3PO_4 is a stable inorganic salt and does not decompose nor vaporize under treating conditions, solution loss is negligible.

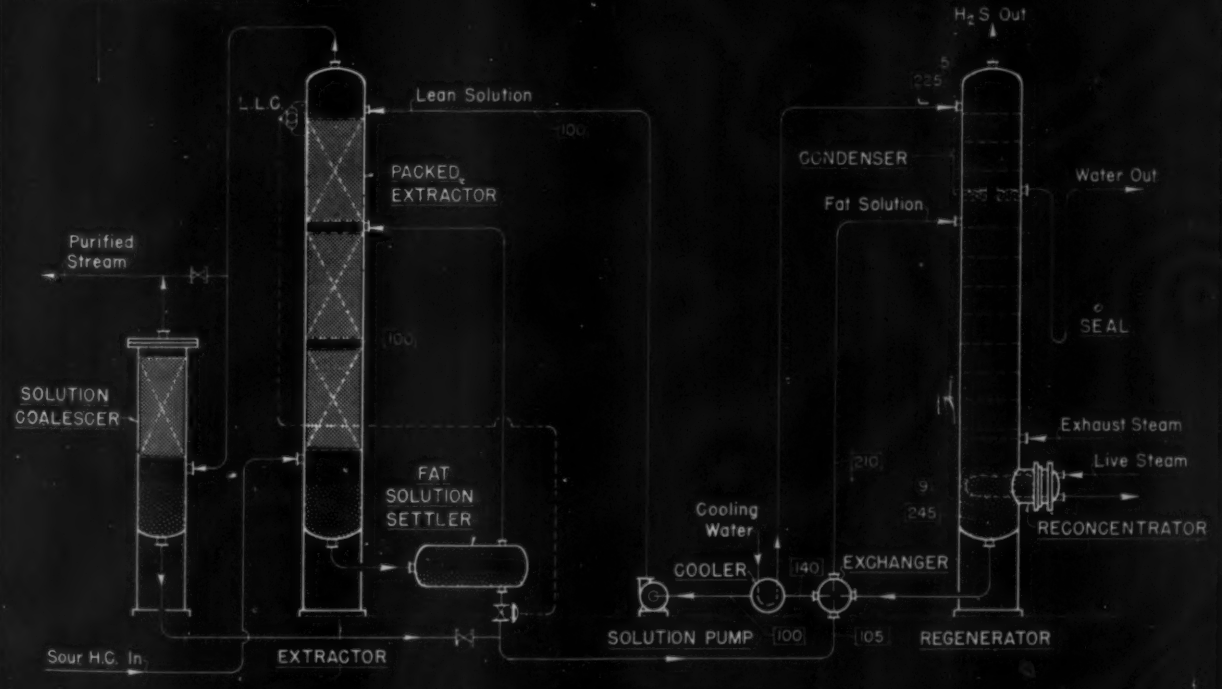
Iron and steel are satisfactory for all equipment. In case of corrosive cooling water, Admiralty-metal tubes and Muntz-metal sheets and baffles may be used for condenser and solution cooler. Gauges, instruments, valves and pumps should be all iron and steel. Observance of these general specifications will insure freedom from corrosion on the process side.

This process is licensed by Shell Development Company, New York.



VAPOR PHASE – INDIRECT REGENERATION

Fig. 1



LIQUID PHASE – DIRECT REGENERATION

Fig. 2

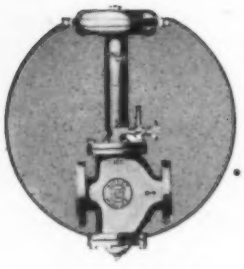
PHOSPHATE DESULPHURIZATION PROCESS

COURTESY. SHELL DEVELOPMENT COMPANY

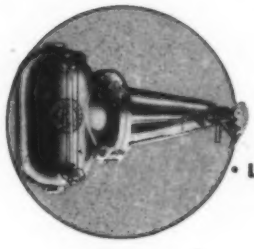
Better Control Valves

OPEN NEW ROADS TO

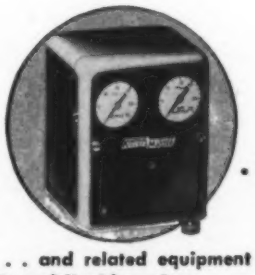
Better Processing



• Diaphragm Valves



• Liquid Level Controls



• Pressure Pilots

. . . and related equipment for steam, gas, air and liquid services.

● With the *new* K & M regulators the petroleum industry gets a green light to more accurate, trouble-free control of pressures, flows, temperatures, and liquid level.

The reason is simply this: *K & M engineers* were given a green light when the line was designed. "Forget what's been done in the past," they were told. "Produce something better . . . to meet the petroleum industry's needs *today*."

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KONTROL MOTOR
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NORTH BERGEN N.J.

Light Hydrocarbon Recovery And Fractionation Processes

FROM humble beginnings in the first decade of this century, natural gas-gasoline processing has progressed at a rapid pace, so rapidly in fact that at the present time, more than 10 percent of the total motor fuel used in the United States is gasoline produced from natural gas. In view of the expansion program now under way, the largest in the history of the industry, this percentage, which even now is an important factor in the nation's petroleum economy, appears certain to increase substantially. It is estimated that the yearly production of natural gasoline and liquified petroleum products will reach 180,000,000 barrels in 1949 when current construction is completed. This is a 45 percent increase over 1947.

The expansion of the industry is attributed to the greatly increased demand for natural gas and liquefied petroleum products, which are being used for domestic and industrial purposes, chemical manufacturing (which currently is consuming approximately one fourth of the LPG production of refineries and natural gas-gasoline plants), blending with gasoline, carbon black manufacture, gas manufacturing, gas enriching purposes, and many other uses.

Plants currently under construction for the recovery of gasoline and liquefied petroleum gases from natural gas generally are designed for the highest propane recovery economically feasible. They employ absorption, refrigeration, and elaborate fractionation arrangements in combinations designed to obtain high and efficient product extraction.

The oil absorption process using spray absorbers was introduced to the industry about 1913, following the compression process developed several years earlier, when a plant was built in West Virginia. A second plant was built in Oklahoma about the same time. Fractionation came into use in the early twenties as a simple economical method of solving the industry's problem of producing a highly volatile, but stable product by making the necessary sharp separation between gasoline and lighter components. After the general adoption of the Thompson-Rafferty type stabilizer, which maintained liquid levels about bubble caps on tower trays for liquid vapor contacting, the use of bubble trays was extended to absorbers, strippers, and other tower applications, to replace the packed towers and spray towers then in general use. A great amount of development work has been accomplished since the early days of natural gasoline recovery. Typical modern plants tend to follow basic patterns, but are actually fairly complex arrangements of a number of carefully worked out processing ideas that are selected or devised to fit the problem at hand.

The methods developed for the recovery of gasoline from natural gas were to some degree adapted to refinery vapor recovery applications in the period between 1915 and 1920. While in early refining operations the recovery of light gases and vapors received little attention, the widespread use of thermal cracking processes and the resulting high gas yields made vapor recovery an important matter. The methods used in the natural gasoline plants were directly applicable.

Today, with gas conversion processes playing a major role in refining operations, and with light hydrocarbons the principal feed and intermediate stock in our ever expanding petrochemical plants, the methods and equipment used for absorbing, stripping, and fractionating these materials constitute an important proportion of all oil industry process plant activities.

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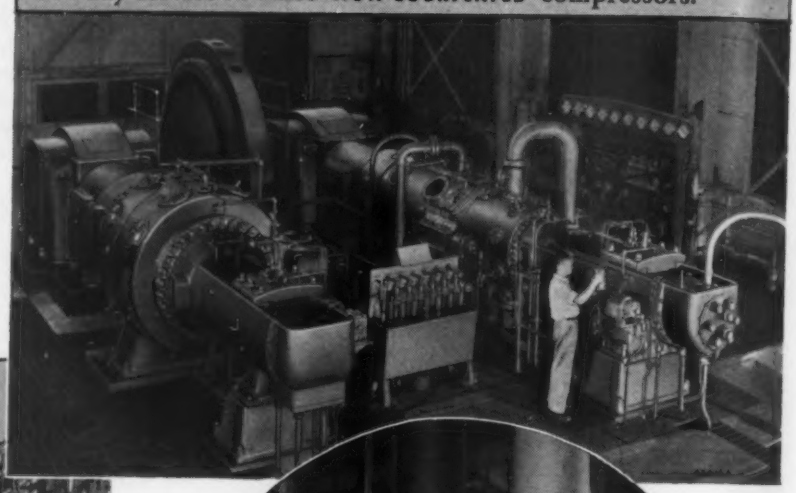

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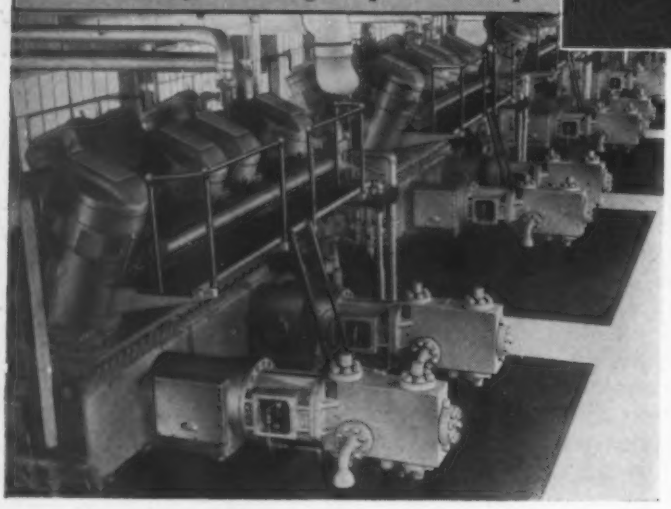

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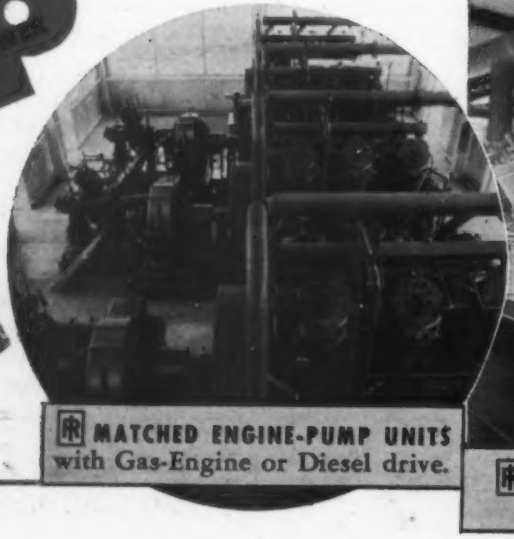

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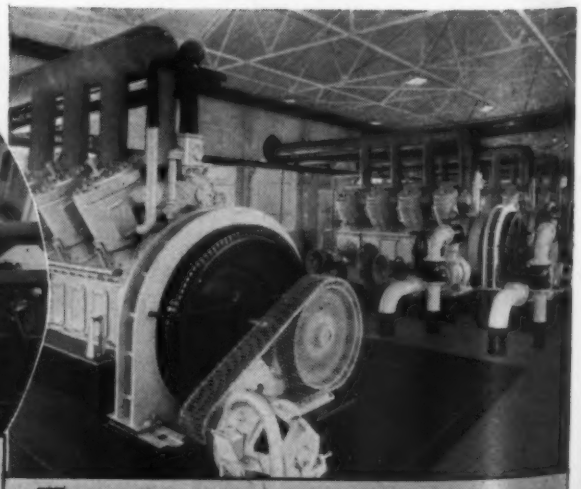

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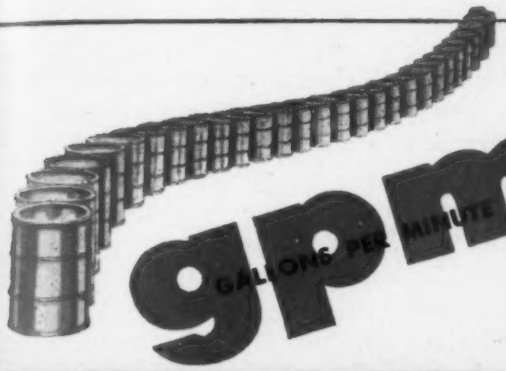

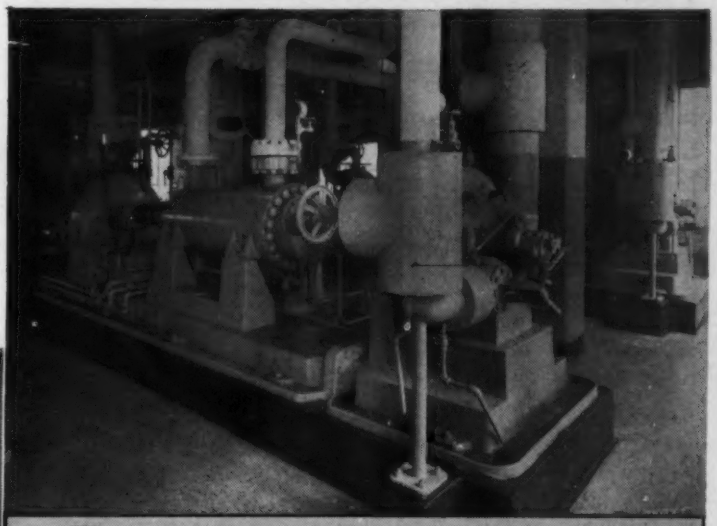

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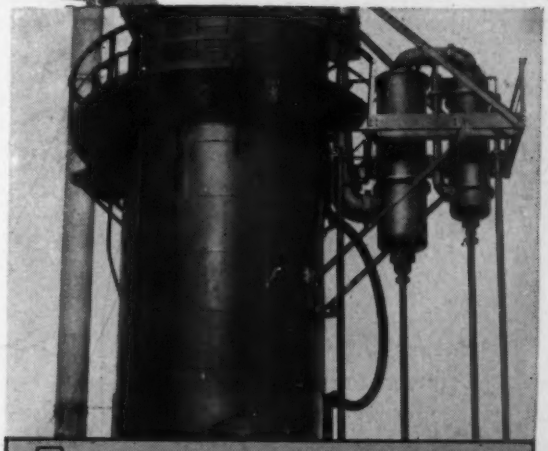

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Low Temperature Gas Rectification Process

PETROLEUM ENGINEERING, INC.

COMPRESSION and refrigeration as a means of separating gasoline from natural gas has a venerable past. As one of the first processes used by the industry, it was in good repute until the expanding market for natural gasoline and the operation of long distance transmission lines gave rise to a demand for a more efficient method, in both extraction efficiency and the possibility of operating at lower pressures. This demand was met by the development of the absorption process.

The original recovery process, that of compression and cooling, with separation by gravity in a common receiver for the cooled gas and condensed gasoline fractions, had the inherent limitation of separation by simple condensation. Under these conditions, the total condensed liquid is in equilibrium with the total non-condensed gas at the temperature and pressure prevailing within the condenser. The result was a condensate containing a large proportion of undesirable light hydrocarbons, and a residue containing a large proportion of the heavier hydrocarbons desirable as gasoline fractions. If the condensate only is further processed, as in the older plants, the desirable hydrocarbons in vapor are lost. Several years ago the gas rectification process was introduced to eliminate these losses. Originally simple water cooling for reflux condensation was used.

Later with the improved market for butanes, the process was improved with refrigeration to obtain practically complete butane recovery. Then the demand for propane resulted in still lower temperatures being desirable and dehydration and two level refrigeration was added.

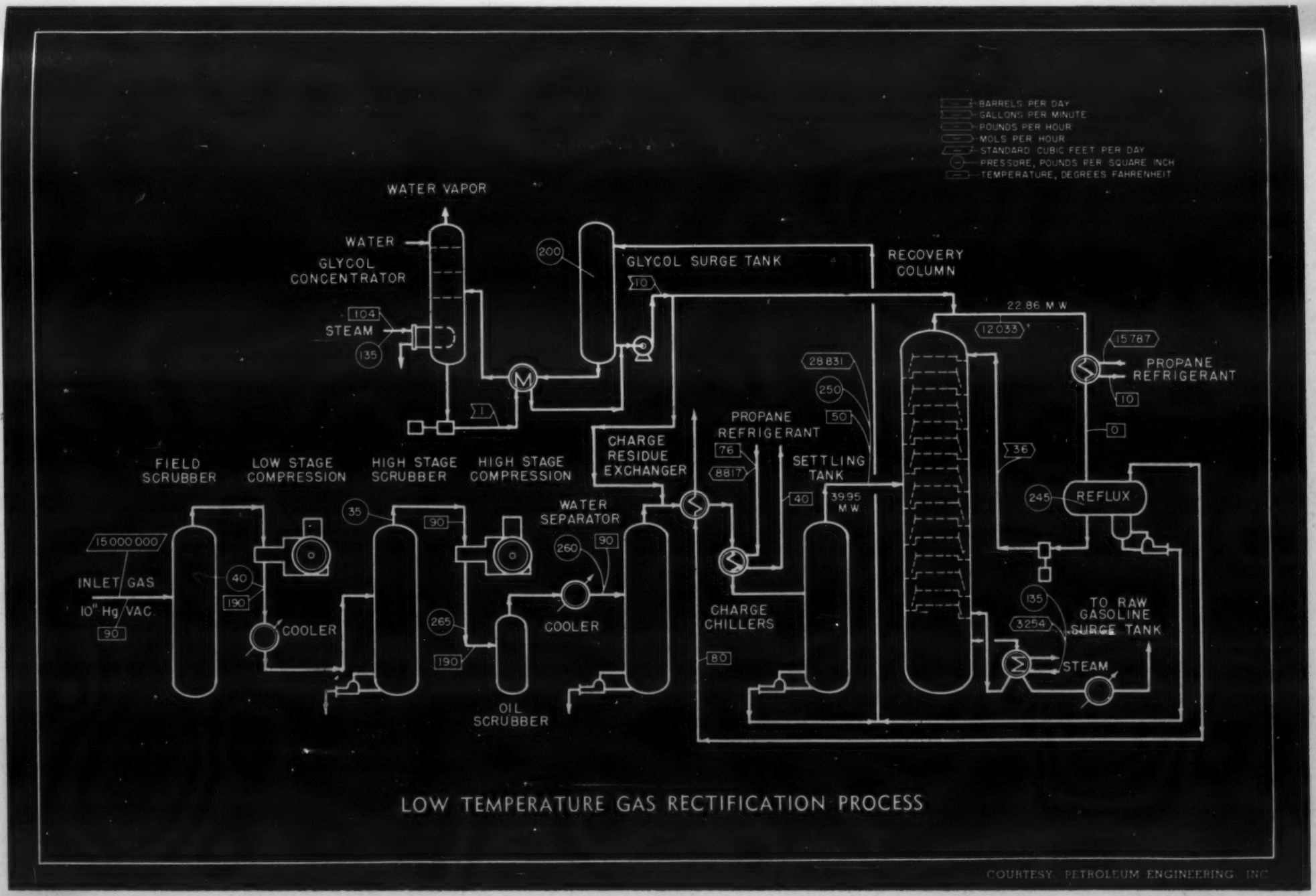
The flow diagram shows a typical low temperature gas rectification gasoline plant. Only the compression-refrigeration and recovery column section is shown, because the fractionation channel is conventional in refineries and gasoline plants.

As in the compression and cooling process, gas from the field scrubbers is compressed and cooled; two stages of compression being utilized in this plant. After the second stage of compression and cooling and after passing through a water separator, a sufficient amount of glycol is added to the stream to depress the freezing point of the water vapor present. Then the gas flows through an exchanger where it receives a preliminary cooling by the cold residue gas from the recovery column reflux accumulator. A propane refrigerant exchanger then cools the gas to 50° F., the condensed water and the glycol are removed in a settling tank and the gas and condensate charged to the recovery column.

In the column, a down-flowing stream of reflux, that has been condensed by propane refrigeration, serves to wash back the desired heavier hydrocarbons, which are carried downward to a point below the feed plate. Below the feed plate, heat generated in the reboiler serves to strip the resulting condensate of excess light components retained by the reflux stream, so that in effect the column operates as a deethanizer. The resulting bottom product of raw gasoline contains practically all the butanes in the gas and a high percentage of the propanes. This product is released to the raw gasoline surge tank for subsequent fractionation.

As shown on the flow diagram, the top temperature of the column is 40° F. This temperature is obtained when overhead vapors from the recovery column pass through an exchanger with propane-refrigerant, which cools the gas to 0° F. to provide condensation for reflux. Although substantially all the water vapor has been removed previously from the main gas stream, an additional quantity of glycol is injected ahead of the refrigerated reflux condenser to effect further removal of water vapor and to depress the freezing point of any water vapor remaining. The glycol and water is removed from the bottom of the reflux accumulator and returned to the glycol concentration unit.

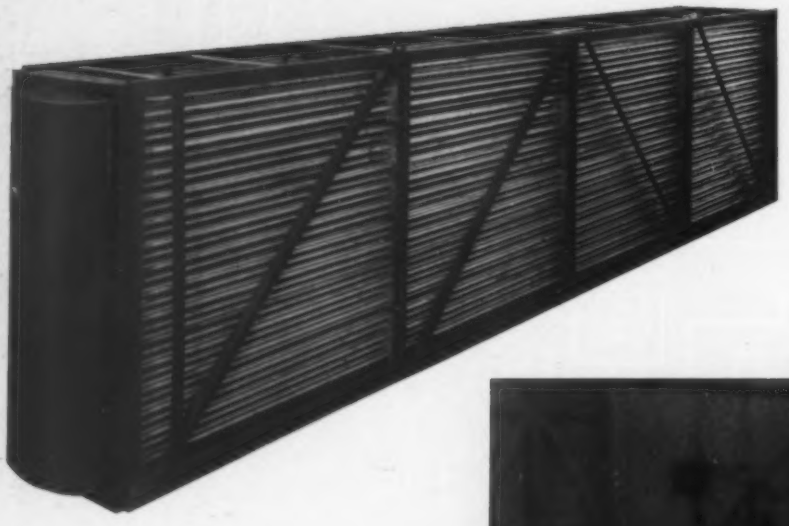
Where the gas is to be compressed for pipe line delivery, or other reasons, to a pressure sufficient for the operation of this process, the cost of the recovery system is comparatively low. The utilities, heating and cooling, are especially low. Compared to the absorption process, fuel and cooling loads are approximately one-half.



LOW TEMPERATURE GAS RECTIFICATION PROCESS

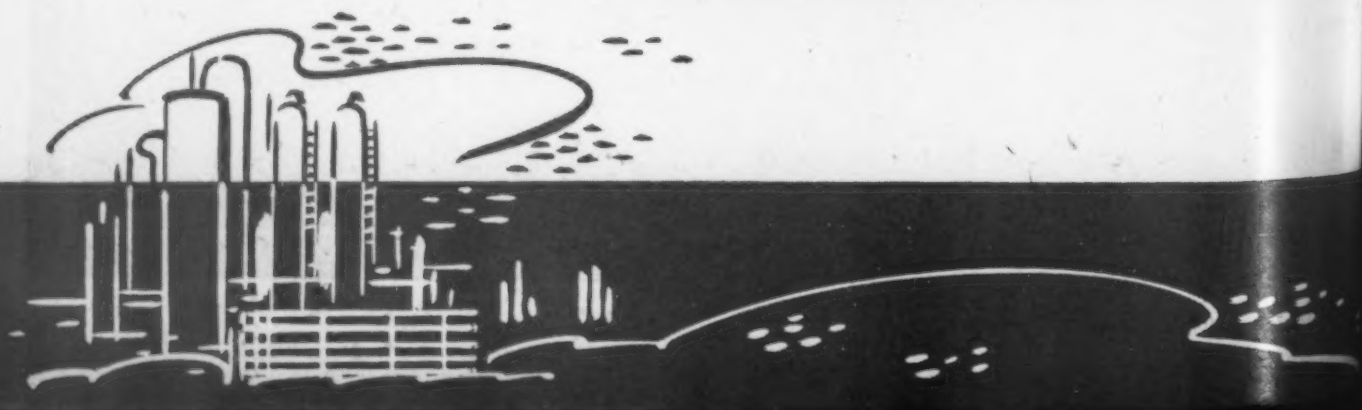
COURTESY PETROLEUM ENGINEERING INC

The MEN, MACHINERY, MATERIALS *behind* WESTERN HEAT TRANSFER EQUIPMENT



Western's vertical atmospheric section with fabricated head, pictured at left, illustrates the close consideration of mechanical detail which has furnished economical solutions to the heat transfer problems of our customers.

To assure consistent design quality, Western enlists the aid of the Magnaflux Corporation to Stresscoat, and Strain Gage (right) castings to insure adequate safety factors at all critical design points.



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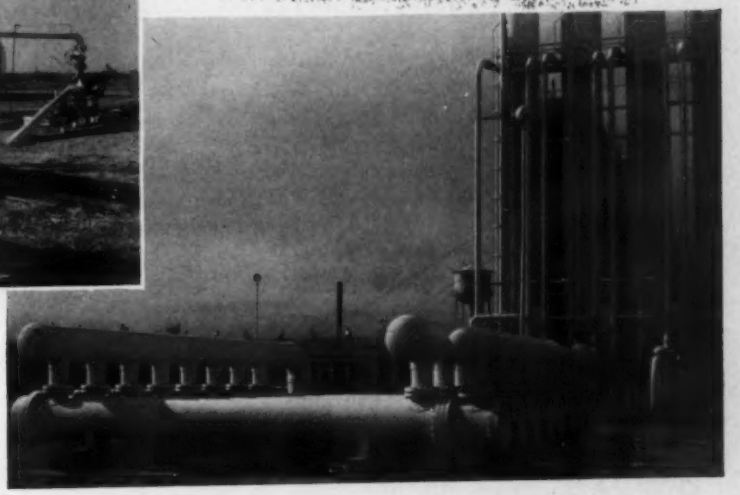

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Absorption Process

FLINT STEEL CORPORATION

A TYPICAL flow diagram of the Tulsa Type Absorption Plant, designed and constructed by Flint Steel Corporation, is shown on the opposite page.

The inlet scrubber is equipped with a single mist extractor and both the absorber and the residue scrubber have double mist extractors to prevent entrainment of mineral seal oil into the residue gas line.

The oil heat exchangers shown on the diagram are of the multitubular, fixed tube-sheet design; however, when it is necessary to clean both through and around the tubes, floating-head removable tube-bundle exchangers are installed.

The evaporator-dephlegmator shown is of the combined type; however, in some instances, the evaporator and the dephlegmator are separate towers. The evaporator is furnished with two internal heating elements, one above each of the two lower trays, in which live steam is used as a heating medium through the tubes. This is a patented feature, and reduces the amount of steam required for heating the oil as there is no radiation loss. The oil passes around the tubes and inside the shells of the internal heating elements.

A small stream of hot mineral seal oil is taken off the bottom of the evaporator, heated with steam and introduced into the oil reclaimer. The oil vapors are condensed and the oil flows into the mineral seal oil storage tank. By continuously reclaiming the oil in this manner, the absorption oil remains in a clean condition.

The vapors from the raw make accumulator tank are recompressed for further processing.

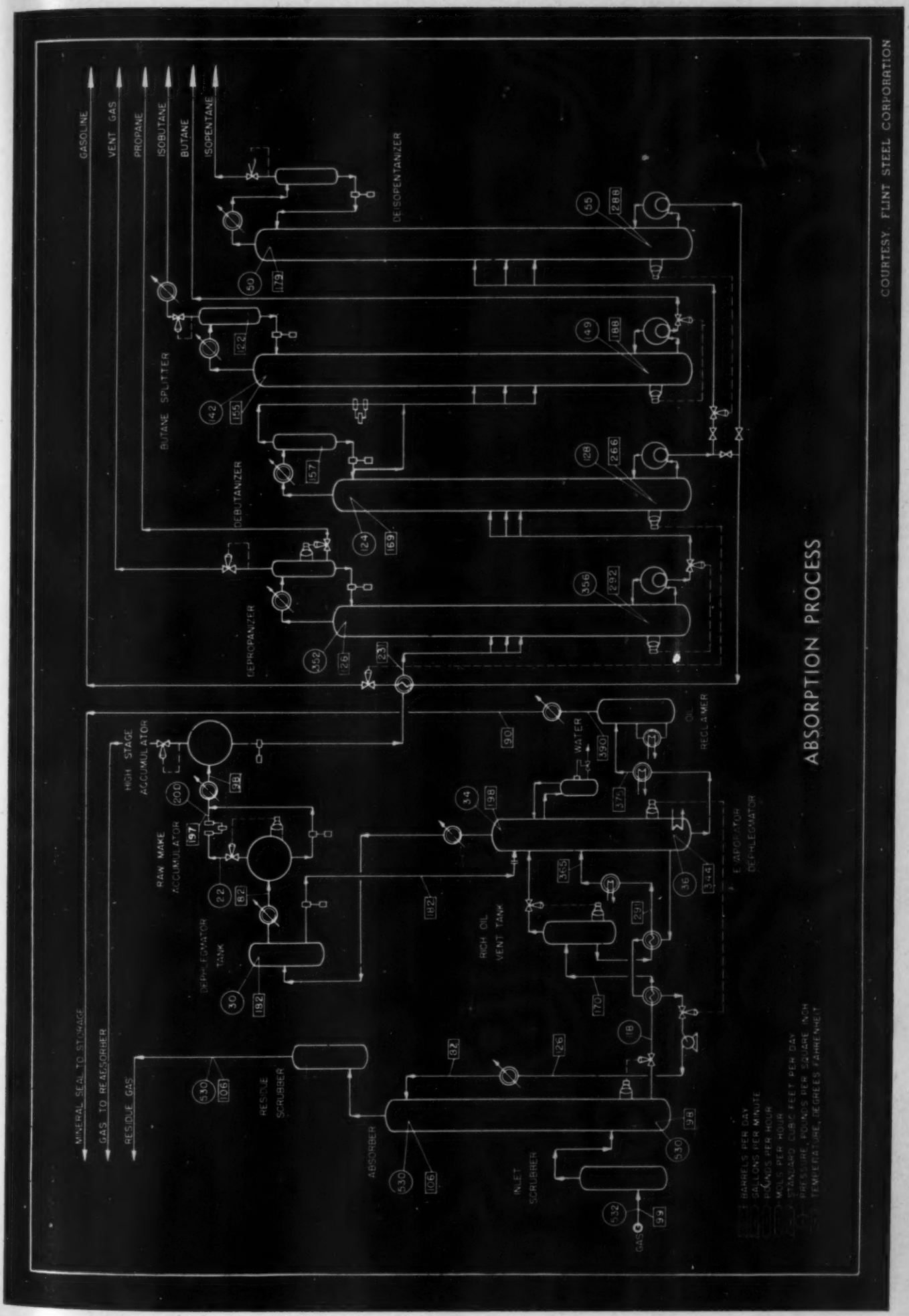
In the fractionator system on the diagram, a depropanizer, debutanizer, butane splitter and a deisopentanizer are shown. In some plants this would be more fractionators than necessary. In some installations, with the absorber working under high pressures, it would be advisable to use a de-ethanizer ahead of the depropanizer in the system.

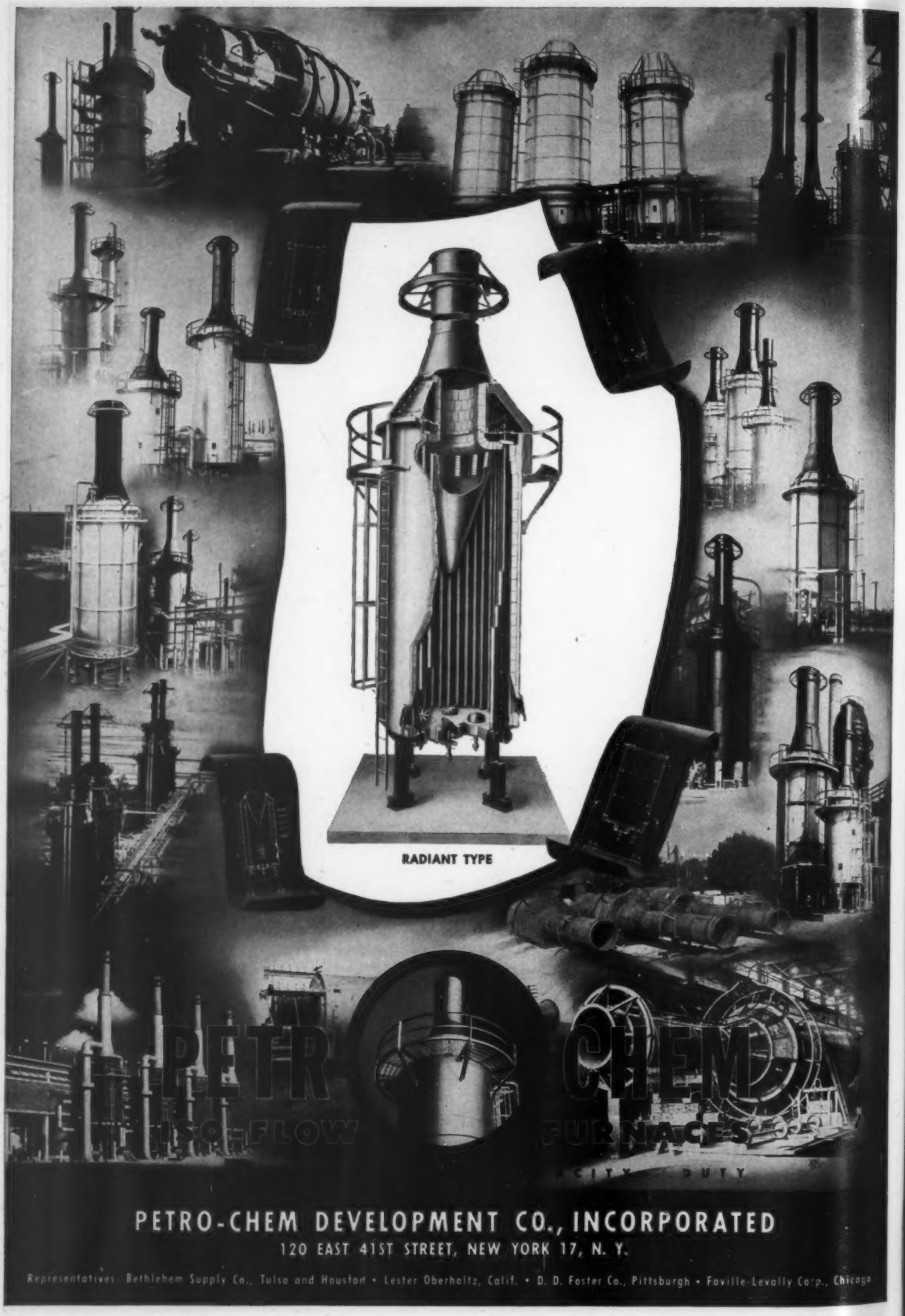
The coolers usually are of the atmospheric type with cast-iron or steel heads and inhibited Admiralty tubes. Some installations require closed type floating-head removable tube bundle coolers. The "Tulsa Type" atmospheric coolers have straight tubes. Both tube sheets are free to move back and forth to take care of expansion and contraction of the tubes. Scale from the tubes is shed by temperature shock which causes elongation of the tubes. Since the tubes are taut and straight, water dripping on the under side of the tubes is uniform; thus, the coolers below the top cooler in each bank receive evenly distributed cooling water. The oil coolers, gasoline condensers, fractionator reflux condensers and fractionator final gasoline cooler are of this type, being built for the pressures necessary.

An absorber usually has 20 bubble trays for relatively high butane extraction. If higher butane extraction or propane extraction are desired, as many as 30 bubble trays are used. On some installations chilling elements are used in the absorber. This is also a patented feature. A refrigerant is introduced through the tubes of the chilling elements and the lean oil passes around the tubes and inside the shells of these internal chilling elements. The bubble trays in the fractionators number from 24 to 50, depending upon the service and degree of fractionation desired.

A gasoline heat exchanger is used on the feed line to the first fractionator for exchanging heat between the feed and the finished gasoline.

Rate-of-flow controls are used to control the amount of reflux to the various fractionators. Controls, regulators, and recording instruments are used throughout the plant to make it automatic in operation. The accompanying flow diagram is typical and can be changed as conditions warrant.


ABSORPTION PROCESS
COURTESY FLINT STEEL CORPORATION
GASOLINE
VENT GAS
PROPANE
ISOBUTANE
BUTANE
ISOPENTANE
DEISOPENTANIZER
BUTANE SPLITTER
DEBUTANIZER
DEPROPANIZER
MINERAL SEAL TO STORAGE
GAS TO REABSORBER
RESIDUE GAS
HIGH STAGE ACCUMULATOR
RAW MAKE ACCUMULATOR
DEPHLEGMATOR TANK
RESIDUE SCRUBBER
ABSORBER
INLET SCRUBBER
RICH OIL VENT TANK
WATER
OIL RECLAIMER
EVAPORATOR DEPHLEGMATOR
GAS
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT



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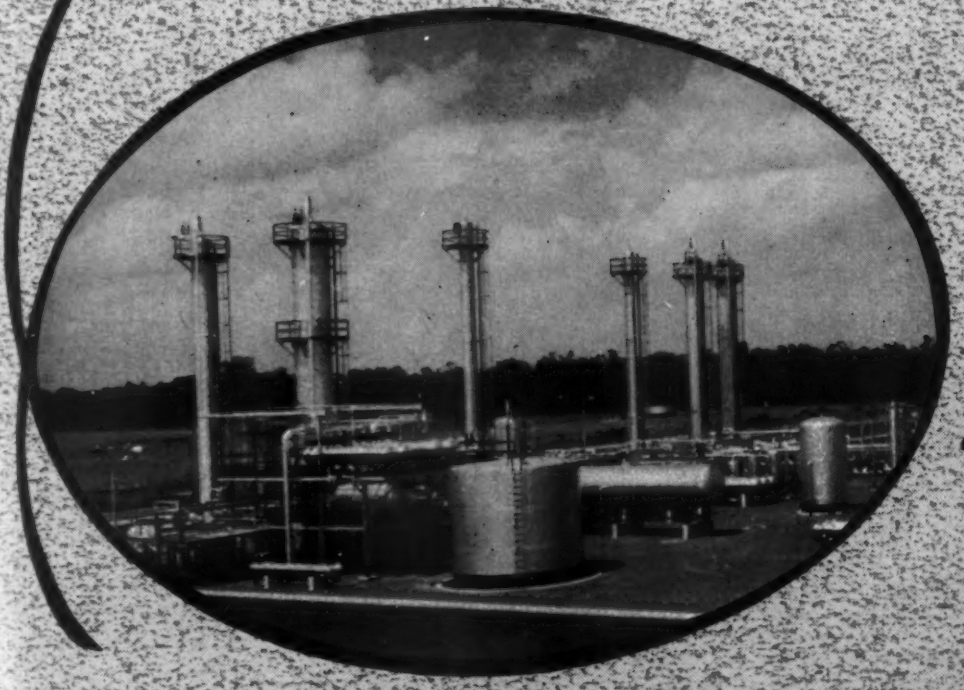


BUILT *Right*

FLINT STEEL CORPORATION

formerly

TULSA BOILER AND MACHINERY CO.

Tulsa, Oklahoma

JONES & LAUGHLIN SUPPLY COMPANY

DURING the last two or three years emphasis in design of gasoline absorption plants has shifted from butane to propane as key component. The market for butanes, propane and L. P. G. has been exceptionally strong.

In setting design conditions and requirements for absorption plants presently under design, it is generally assumed that substantially all butanes can be economically recovered and at the same time a high recovery of propane is assumed. However, such assumptions cannot be safely made without a careful economic study of the particular situation under consideration. Higher propane recovery increases both first cost and operating cost rather sharply. High propane content of gas and high pressure for absorption have much to do with the economic feasibility of high propane recovery.

The flow diagrams on the opposite page represent modern versions of two types of plants, the upper for use where butane is the key component, the lower where propane is the key component. Both plants are flexible so that operation may be varied with market conditions to produce optimum plant returns. Stabilizer sections are omitted from both flow diagrams.

The "Moderate Propane Recovery" plant usually will recover from 25 to 35 percent of the propane. The limiting factor is the still pressure, which must be kept low enough to permit stripping at the available hot-oil temperature without using uneconomical stripping steam quantities. The use of a direct fired oil preheater as shown permits higher still pressures, and, consequently, higher propane recovery.

Between the absorber and the still, a fat-oil stabilizer (or fat-oil deethanizer) is shown. Heat from normal oil-to-oil exchange is applied to the feed entering this unit and, consequently, better heat economy is secured than when heat other than from oil exchangers is applied to the stream before leaving the bottom of the fat-oil stabilizer. By controlling preheat of the lower charge stream and regulating cold oil quantities to the top and middle sections, substantially all the methane and 60 to 80 percent of the ethane can be removed from the fat-oil. This is accomplished without excessive losses of propane and heavier. Fat-oil stabilization reduces the quantity of recycle vapors from the still and stabilizers with resultant savings in equipment size and cost, while at the same time overall butane-propane recovery is increased.

It will be noted that a lean oil absorption section is set on top of the fat oil stabilizer proper. This may be omitted and the overhead vapors compressed to enter the main absorber. In many cases this operates to better advantage. In connection with this "Moderate Propane Recovery System," the absorber pressure is assumed to be 200 psig or more.

The "High Propane Recovery" plant indicated by the lower flow diagram is satisfactory for propane recoveries on the order of 50 to 85 percent. This high propane recovery is economically feasible only when gas from the field is relatively rich in propane and when absorber pressures of from 400 to 700 psig are available.

Distinguishing features of this system are the high- and low-pressure stills, and the reabsorber where the cold fat-oil serves as a large portion of the absorption medium required for the reabsorption of the heavier constituents of the relatively high quantities of recycle gases. The primary still is operated at about 225 pounds pressure in order to condense a high proportion of propane in the stabilizer-charge liquid. Since satisfactory stripping is impractical at this pressure, the primary still bottoms are fed to the secondary still, which is operated total condensing at 50 to 75 pounds pressure with the net overhead product being pumped to an intermediate point in the primary still condenser. The addition of this secondary still product aids in holding the maximum amount of propane in the liquid state.

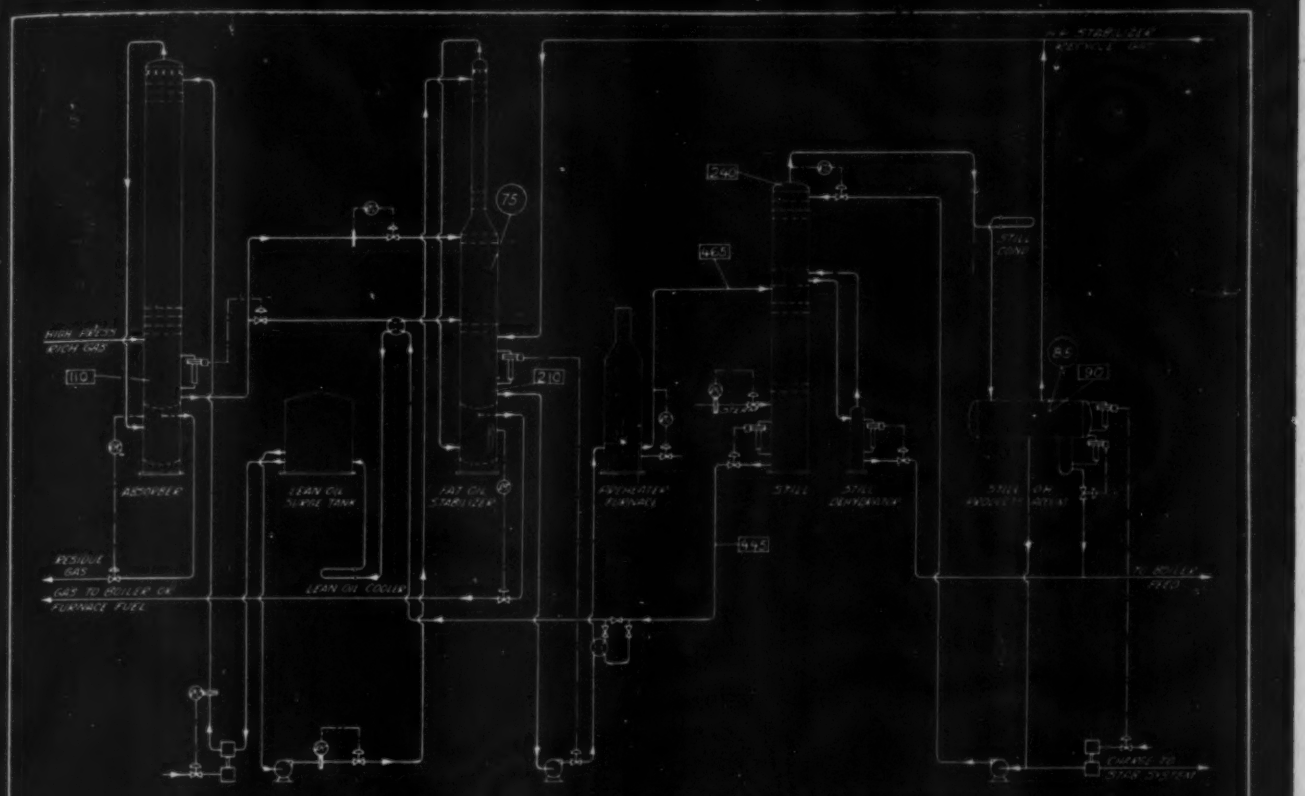
In many cases the lean oil absorber shown as the top section of the reabsorber can be dispensed with and a recompressor substituted for the purpose of compressing this recycle gas to the main absorber inlet.

While not indicated on the flow diagram, it may be desirable to apply a small amount of heat to the base of the reabsorber to assist in the fractionation between ethane and propane and keep down the recycle gas quantities. The secondary still preheater which is dotted in on the flow diagram, may not be necessary.

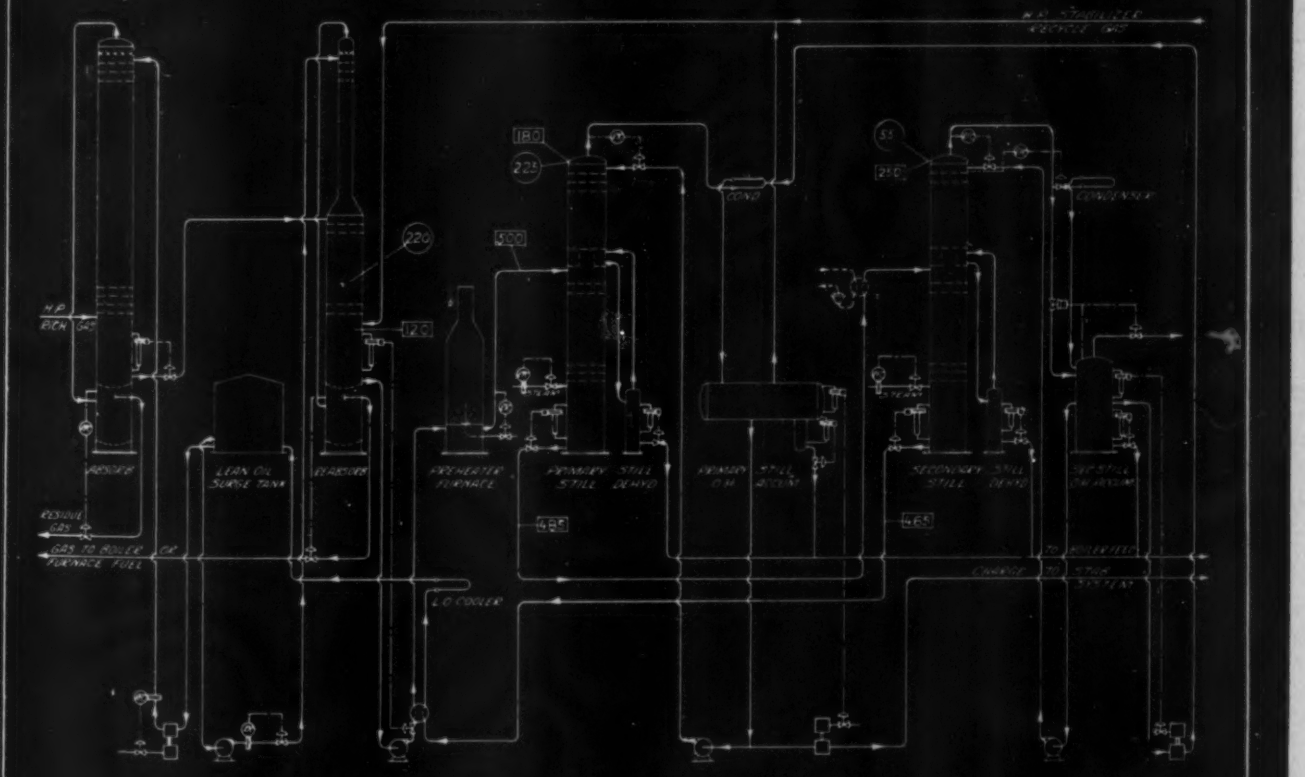
Many variations should be considered before selecting final process arrangement. Among these we would mention principally the application of refrigeration to the absorption cycle and, on occasion, to net overhead vapors from distillation unit, and to net overhead vapors from deethanizer. Also the distillation unit can be reduced to single still operation if recompressors are added for increasing pressure of non-condensable overhead hydrocarbons so that a higher pressure condensing point comparable to primary still condenser can be set up.

As stated at the beginning, each plant design must be carefully studied in relation to its particular operating situation before final decision is reached in regard to economic recovery of propane as well as several other points above mentioned.

Both the flow diagram opposite and this description were provided by Jones & Laughlin Supply Company, Tulsa, Oklahoma.



ABSORPTION SYSTEM FOR MODERATE PROPANE RECOVERY



ABSORPTION SYSTEM FOR HIGH PROPANE RECOVERY

BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

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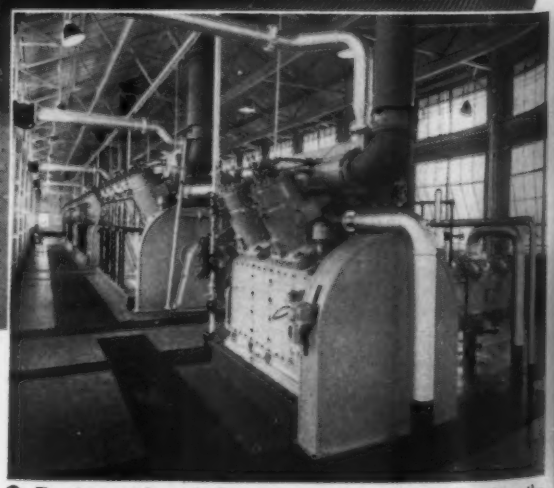
These four space saving GMV's handle 25 million cu. ft. of casinghead gas daily through two stages of compression from 60 to 215 psig. and from 210 to 800 psig.



● Engine side of the compressor line, showing the 200 hp Cooper-Bessemer GMX in the foreground, used to re-compress still vapors to 300 psig.

ANGLE

Compressors by

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Cooper-Bessemer powers Humble's Clear Lake Gasoline Plant

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50 MILLION CU. FT. OF GAS DAILY
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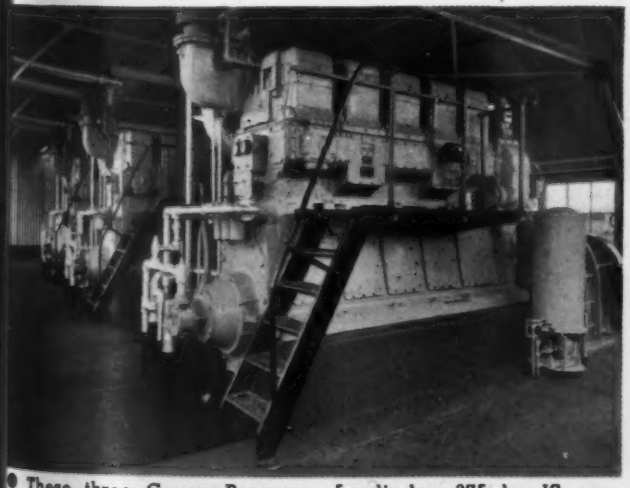
Here, 45,000 gallons of de-ethanized products, including gasoline, butanes, and propane are produced daily. The largest engine-driven compressors, the 10-cylinder GMV's totalling 4,000 hp, handle 25 million cu. ft. of casinghead gas a day, through two stages, from 60 to 800 psig. An additional 19 million cu. ft. of gas well gas, plus 6 million cu. ft. of high pressure casinghead gas delivered from another source, makes the total 50 million feet processed daily.

The Cooper-Bessemer GMV has long proved to be the industry's top performer for all kinds of compressor service within its 400 to 1000 hp. range. There's good reason for its widespread use. Check up. It can save you money on space and housing, on maintenance, and on over-*all* operating cost!



● These three Cooper-Bessemer 5-cylinder, 375 hp JS gas engines drive 250-kw, 440-volt, 60-cycle generators to supply plant and camp electric power.

● Three 140 hp Cooper-Bessemer GAW horizontal gas engines direct-driving high pressure lean oil pumps for supplying absorption oil to the plant's two absorbers at 800 psig.

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Stage Separation Cycling Process

O. L. OLSEN

STAGE separation plants for recovery of liquid hydrocarbons from natural gas have been found advantageous in small condensate fields, or where installation of a full-scale absorption type plant is not feasible. The low installation and operating costs of stage separation plants often permit production of condensate fields that otherwise would be shut-in.

The stage separation process employs comparatively simple equipment, and is well-adapted to high reservoir pressures and the processing of small volumes of gas, because its operation employs pressure reduction in each stage of separation. Such plants operating at high pressures seldom utilize more than three separation stages, and compressors are necessary only for compressing residue gas for field injection, or pipeline sales.

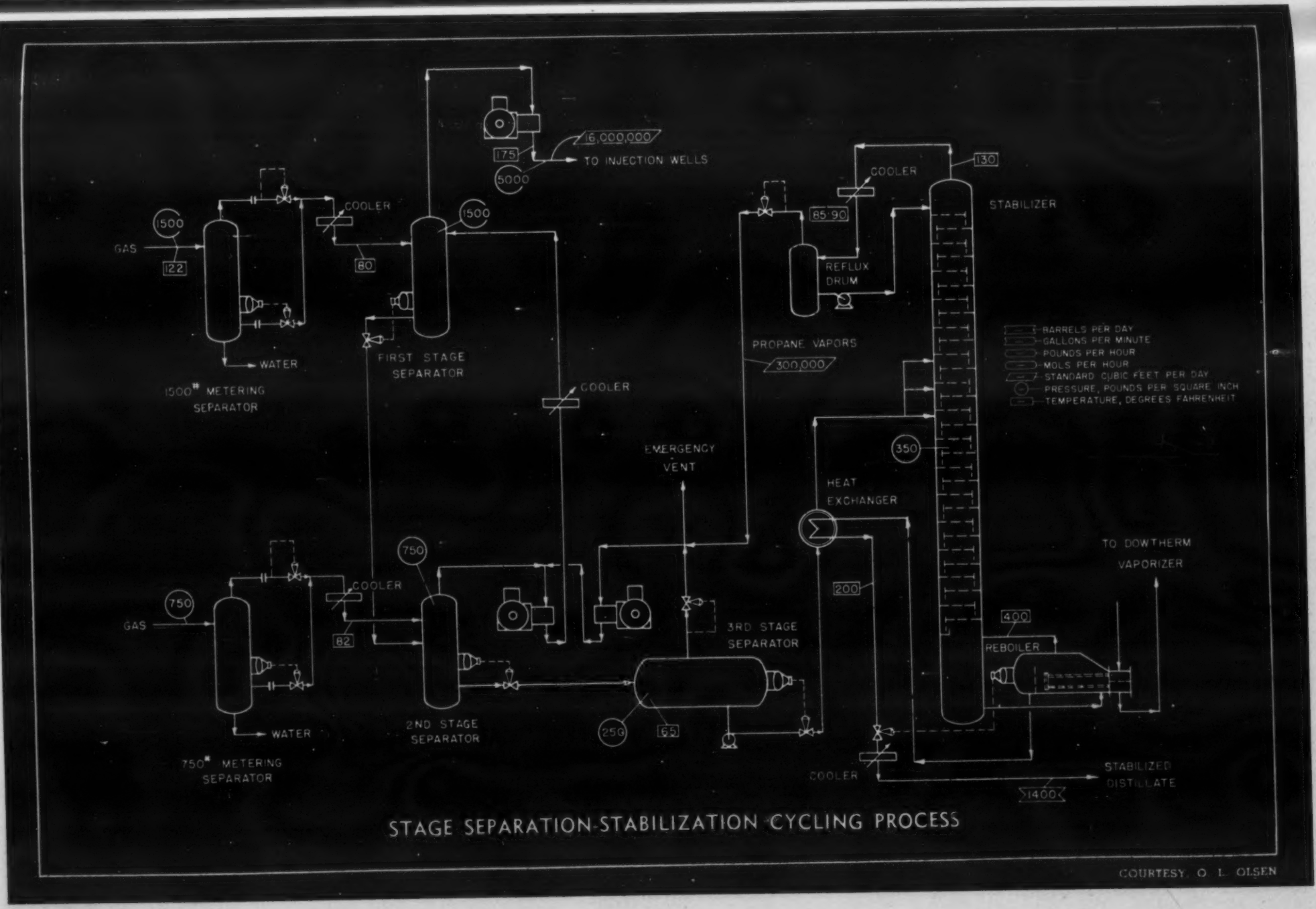
In the accompanying flow diagram of a three-stage recovery plant, the flexibility of such processing is demonstrated in the handling of two separate streams of gas from production at two different pressures. Although not shown, the plant actually utilizes five separators for metering purposes—two operating at 1500 psig and two at 750 psig, and a fifth separator for use as a test separator. After metering, the gas and liquid are recombined before entering the process stage separation system.

There are three process separators—one operating at 1500 psig, one at 750 psig, and a final separation of liquid hydrocarbons from the gas in a flash tank operating at 250 psig.

Combined gases and liquids leaving the 1500 psig metering separators are cooled in an atmospheric section and enter the first stage 1500 psig process separator. Liquid condensate is collected in this separator and led under level control to the second stage 750 psig process separator. Combined gases and liquids from the 750 psig metering separators flow to the second stage process separator, also, passing through a cooler to join in an equilibrium flash with the bottoms from the first stage 1500 psig process separator. Liquid bottoms from this separator are flashed into a third stage 250 psig process separator. Vapor from this third stage separator is combined with vapor from the stabilizer reflux drum and then recompressed to the 750 psig level, combined with the vapor from the 750 psig separator and recompressed to 1500 psig. The gas is cooled in an atmospheric section after compression and returned to the 1500 psig separator. All gas overhead from the 1500 psig separator is picked up by the compressors, boosted to 5000 psig and returned to the field injection wells.

Raw distillate from the final flash tank is pumped through heat exchange with the final product and charged to the stabilizer, a 30-tray column, at 350 psig and about 175° F. Sufficient overhead vapors are condensed at 85-90° F. to supply reflux and the uncondensed overhead vapors are returned to the compressors for field injection. Reflux is 30 gallons per minute, and the volume of uncondensed propane-rich vapor is 300,000 cubic feet daily. Stabilized distillate is produced at the rate of 1000 barrels per day from the bottom of the stabilizer. Heat to the stabilizer column reboiler is supplied by a Dowtherm unit, instead of steam, because the reboiler is the only item of equipment in the plant requiring heat and the installation and operation costs of this unit are less than that required for a steam plant. The two pumps, one pumping feed to the stabilizer and the other pumping reflux on the tower, are driven by electric motors on current supplied by the plant's generator units.

In the plant discussed herein, two 800-horsepower compressors are used to compress the 16,000,000 cubic feet daily of gas through three stages to the formation injection pressure of 5000 psig. Each producing well and injection well is provided with a separate line and production and injection rates controlled by adjustable chokes in the plant. Each line is equipped with flow meter, presure gauge and thermometer.



STAGE SEPARATION-STABILIZATION CYCLING PROCESS

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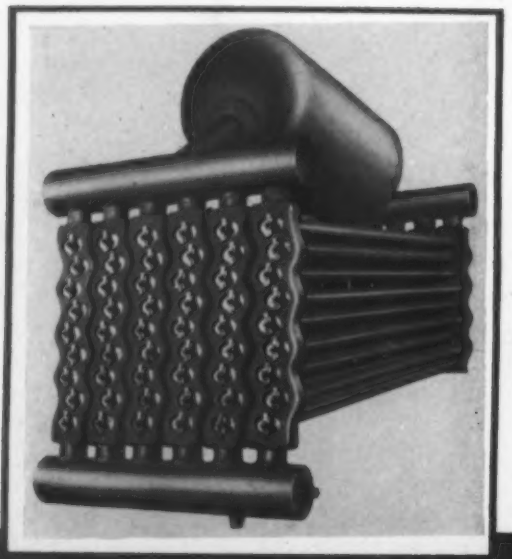

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Absorption Cycling Process

CYCLING is a recovery process in which condensate or distillate is extracted from high pressure natural gas, and the residue gas compressed and returned to the producing formation. The cycling method of production and recovery is applicable to and necessary in reservoirs where the hydrocarbons present exist in a single vapor phase to maintain the original pressure and prevent condensation and resultant loss of liquid hydrocarbons in the reservoir. Cycling provides for maximum recovery of hydrocarbons from the reservoir and the prevention of surface and underground waste of natural gas. A well-planned efficient cycling program operation will often recover an extremely high percentage of the hydrocarbons in the reservoir compared to production without cycling. Also, the dry gas returned to the formation is readily available and may be withdrawn at a later date for use in any advantageous manner in addition to its immediate utility for pressure maintenance.

Varying plant designs are used for cycling operations. Simple separation installations, which were once quite widely used, are based on the principles of retrograde condensation and extract the condensate by a reduction in pressure. A variation of the simple retrograde condensation plant required cooling of the gases, since in the retrograde range condensation at a given pressure increases at lower temperatures. Extraction efficiencies were thus improved by the addition of a refrigeration cycle to reduce the temperature of the wet gas before separation.

Possibly the most effective type of cycling process, which is discussed here, utilizes high pressure absorption to recover the condensate. This type of plant has a much higher extraction efficiency than those processes mentioned above. Also, in addition to higher extraction efficiency, less horsepower is required for compression and injection of the residue gas back into the formation, due to the higher suction pressures employed. A large majority of the cycling plants constructed during the last few years employ the absorption process. The plant investment, however, is large with the absorption process, so that usually large volumes of gases must be available for processing and adequate reserves are necessary to justify the construction of a plant.

Gas wells supplying cycling plants have high delivery pressures, usually in the range of 1500 to 5000 psig. The residue gas must be returned to the producing formation. It is processed and then compressed to a sufficient level over reservoir pressure to allow injection into the reservoir through special wells set aside for the purpose. The necessary injection pressures are high and the attendant equipment and operating expenses are major factors in plant design. Because of these factors, the main plant absorbers are operated at quite high pressures to avoid increased recompression expenses. Absorbers commonly operate in the range between 1000 and 2000 psig., with 1500 to 1800 psig. perhaps most frequently used. An attempt is made to achieve a plant design having optimum economic balance between the factors of recompression costs, operating costs, equipment costs, and products value.

In the plant diagram shown, lean absorption oil of 35-40° API is charged to the absorbers by a multi-stage centrifugal pump. The oil has a molecular weight of 200-225, an initial boiling point of 480-500° F. and a final boiling point of 650° F. The absorption pressure is about 1500 psig., although absorption pressures as high as 2000 psig. are being used in similar plants.

Rich oil from the absorber is flashed in the base of the high pressure reabsorber and some of the light hydrocarbons, methane and ethane absorbed in the oil are released from solution along with heavier components that it is desirable to retain. These vapors are contacted with lean oil to recover the heavier components. A substantial portion of the methane and ethane is released into the residue gas stream to the compressors. The rich oil from the reabsorber is flashed into a lower pressure reabsorber which is constructed in combination with a stripping column. Flashed rich oil flows down over bubble plates in the stripping column to be contacted with a stream of propane rich vapors which tend to strip out any absorbed ethanes and lighter to replace them with propanes and heavier.

Tail gases from the reabsorbers are recompressed by a system of stage compression to the main absorber pressure and join with the absorber residue gas which is raised to field injection pressure level by a final stage of compression.

The combined rich oil streams from the absorber, high pressure reabsorber and low pressure reabsorber are heated by exchange with the lean oil then in a direct-fired heater and charged to the still. In this tower, the lean oil is stripped and removed from the bottom of the tower and the raw distillate is taken overhead. The distillate is condensed for reflux to the still and for charging to the stabilizer. Uncondensed vapor from the reflux accumulator is returned to the low pressure reabsorber for recovery of propanes. Raw distillate charged to the stabilizer is split into stabilized natural gasoline as a bottom product and the propane and butane overhead receives final fractionation in a deethanizer. Overhead vapors from the deethanizer are used for fuel.

ABSORPTION CYCLING PROCESS
FUEL GAS
DEETHANIZER
EXHAUST STEAM
LPG PRODUCT
STABILIZED GASOLINE
STABILIZER
STEAM
FIRED HEATER
STRIPPING STILL
L.P. REABSORBER
H.P. REABSORBER
MAIN ABSORBER
GAS TO FIELD INJECTION
GAS FROM FIELD
T BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MILS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT

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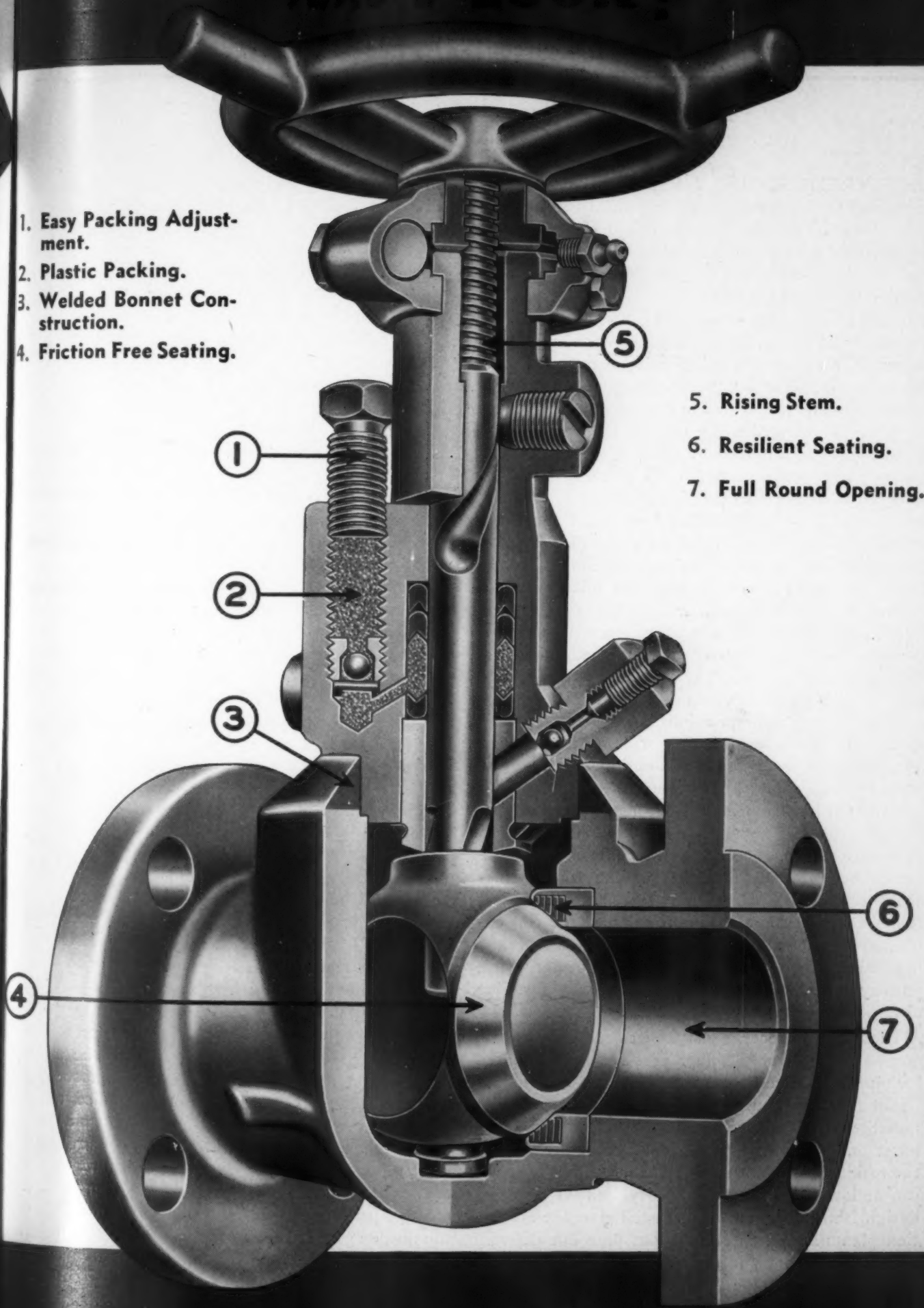

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Pipeline Absorption Process

THE TREMENDOUSLY increased consumption of natural gas as domestic and industrial fuel and as a raw material for the manufacture of chemicals by various synthesis processing plants has resulted in the construction of many long distance high pressure gas transmission pipe lines.

In general, the gases handled by these lines are of necessity rather dry, since wet gases would result in excessive condensation of heavier components in transmission and at compressor stations. The gases, too, are usually dehydrated to low moisture contents. Since field gases ordinarily contain recoverable hydrocarbons, gases moved by pipeline commonly are processed for gasoline and LPG extraction in the field before delivery to the line. Alternately, the recovery plant may be located at some point along the length of the pipeline as conditions may dictate.

In many cases gases for sales to pipelines are collected at low pressures from crude oil separation systems or other field sources. In such instances stage compression methods are usually necessary to raise the gases to pipeline delivery pressure which is often several hundred pounds. The attendant natural gasoline recovery plants are then designed to take advantage of the stage compression and condensation effects and ordinarily utilize absorption at near pipe line delivery pressures to obtain optimum recovery of propane and heavier hydrocarbons.

Undoubtedly the most efficient installations of pipe line absorption plants are those made where large volumes of gas are produced at sufficient pressure for delivery directly into the pipeline without mechanical compression. Where extremely large gas reservoirs exist some of the production problems encountered in smaller fields are less serious, or less urgent. Cycling in many such fields is not necessary, or may offer little ultimate gains. Many pipe line absorption plants are so situated.

These plants receive gas feed at field conditions, often at considerably higher pressure than required for pipe line delivery. The absorber pressure thus can be set at the optimum point in the available range to give the most economic product recovery and equipment and operating costs. Since recompression of the residue gas is not required the absorber pressure can be lowered to any desired level as set by other design factors. Actually, most such pipe line absorption plants are operated at about pipe line delivery pressure, frequently in the range of 700 to 1000 psig.

In many cases pipe line absorption plants receive comparatively lean gases. The fact that such plants often can be positioned to handle large volumes of this lean gas, perhaps as much as 100,000,000 to 250,000,000 cubic feet daily, is quite important to the economics of these operations. The extreme simplification of such plants through the avoidance of compressor loads is also an important factor. Some plants have little or no compressor load. The amount of gas which is released through the reabsorbers and other low pressure vent sources is often just about equal to plant fuel requirements. Any excess is recompressed.

A typical high pressure pipe line absorption plant is shown on the opposite diagram. Field gases are absorbed in a high pressure absorber, rich oil is flashed in a reabsorber to release ethanes and lighter, the rich oil is stripped in a steam stripping still, product is recovered overhead at the still as a liquid, is stabilized for gasoline production, and stabilizer overhead deethanized in turn for the production of propanes. Overhead noncondensables from the still are routed to a reabsorber. Certain close resemblances between this plant and the usual absorption type cycling plant are clearly evident. There is no basic difference in process flow. Because of the lower main absorber pressure a single reabsorber is practical.

Residue gas from the absorber must usually meet a low specified moisture content on delivery to a pipeline. The dehydration of the residue gas is usually necessary, and the method perhaps in widest use is by contacting with adsorbents in solid bed vessels. The adsorbent vessels are in turn stripped of moisture by a reverse flow of heated dry gas. Two or more adsorbent towers are used on a cyclic operation. Where the field gases as produced contain appreciable amounts of hydrogen sulfide, the solid bed adsorption dehydration method is frequently replaced with liquid absorption methods of dehydration which also effect the desulfurization of the gas.

The recent trend in absorption plant design toward quite high propane recoveries has led to the use of refrigeration for lowering the effective absorption temperatures. When refrigeration is applied to the cooling of the lean oil stream to the top of an absorber the temperature of the residue gas stream is also reduced, and a sharp reduction in moisture content of the gas results. A comparatively small temperature change, perhaps 20° F., may halve the moisture in the residue gas. This effect is especially useful in pipe line absorption plants, since the required capacities of residue gas dehydration plants can be greatly reduced with substantial savings in plant investment.

PIPE LINE ABSORPTION PROCESS
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
GAS FROM FIELD
GAS TO PIPE LINE SALES
WATER SEPARATOR
FUEL GAS
WATER
SOLID BED DEHYDRATORS
GAS TO DEHYDRATION
MAIN ABSORBER
REABSORBER
STRIPPING STILL
FIRED HEATER
STABILIZER
DEETHANIZER
FUEL GAS
STEAM
EXHAUST STEAM
LPG PRODUCT
STABILIZED GASOLINE
DEHYDRATION PREHEATER
GAS

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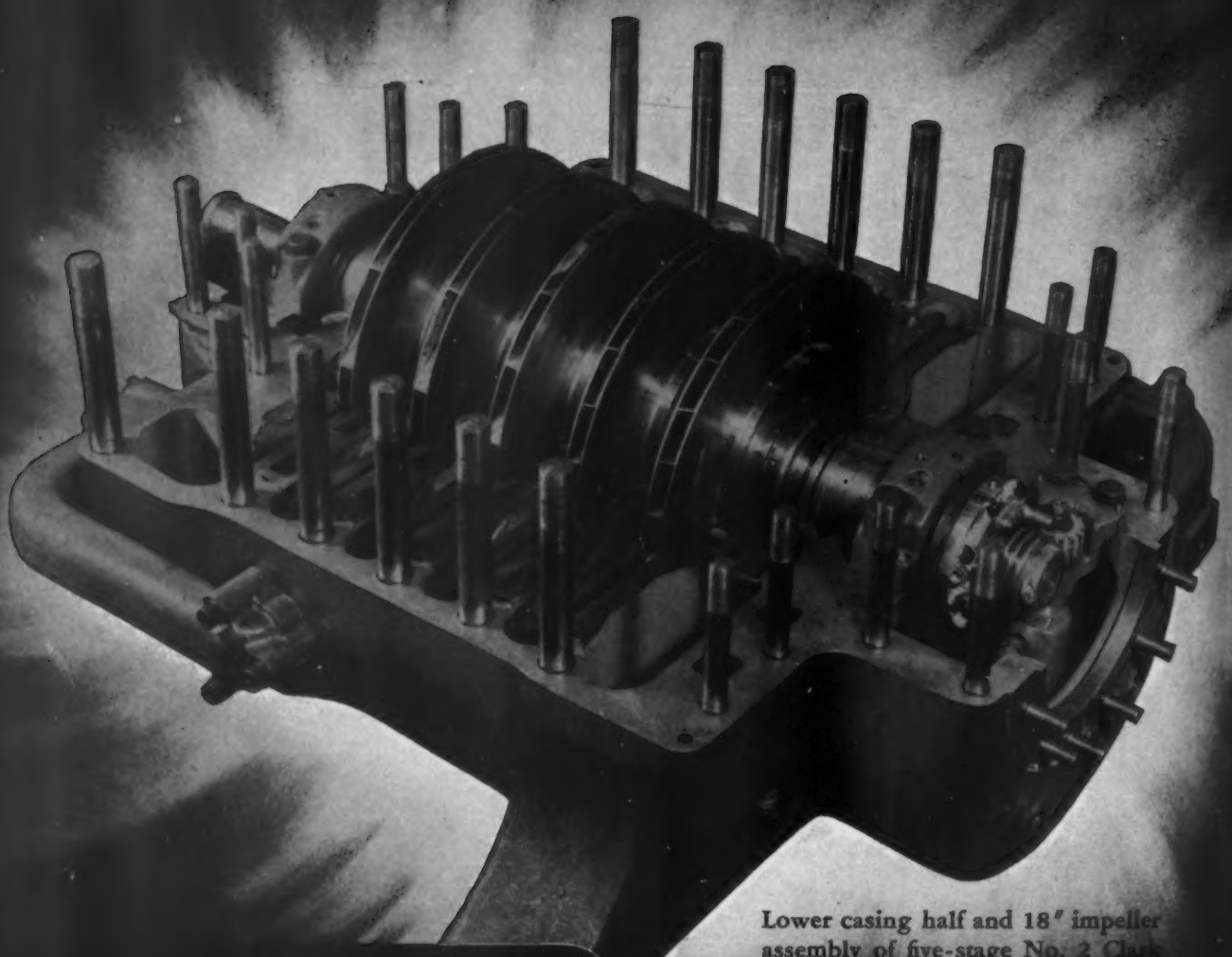

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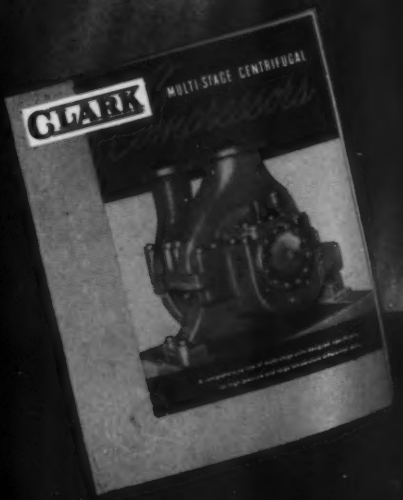

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Absorption and Stripping Processes

ABSORPTION plants are alike in principle but vary widely in equipment and process arrangement. The basic design is shown on diagram 1 opposite. Rich gases are countercurrently contacted with lean absorption oil in a bubble plate tower. Residue dry gas leaves the absorber to sales or other disposal. Enriched oil flows through a preheater into a still where absorbed light hydrocarbons are stripped out and condensed overhead as product. Lean oil is cooled and recycled to the absorber. Condensed product serves as still reflux. Noncondensable gases are routed through a reabsorber or compressed for recycle to the main absorber. Product is stabilized to the desired vapor pressure in a fractionator.

This basic process scheme has been much used. In recent years the increasing use of butanes and lighter hydrocarbons for LPG sales and conversion to motor fuels or petrochemicals, plus the development of high pressure absorbers to process cycled gases from underground reservoirs, has stimulated intense development in design, and flow diagrams have become more complex. Several modifications of the basic plan are shown. It should be noted that some of the ideas are covered by patents which will not be detailed here.

Increased propanes-butanes extraction entails additional absorption of ethanes and lighter components, with adverse effects on stripping still operation. Low pressure at the still is desirable to assist in stripping lean oil and to avoid excessive bottoms temperatures. Low pressures, however, make condensation of overhead vapors more difficult, the condenser temperature being set by the available cooling water.

Absorption temperature is a controlling factor for the extraction obtained. Lower temperatures can be achieved by increased water cooling or by refrigeration. Cooling can be applied to the lean oil, to the absorber through intercoolers, or to the incoming gases as indicated by diagram 2. Lower temperatures improve the selectivity of absorption of heavier components over lighter gases. Since absorbed ethane and lighter gases require additional equipment for their elimination this is important. Lowering the lean oil temperature to the top of the tower also lowers the residue gas temperature, and the moisture content of this gas. For gases flowing to a dehydrator or to pipe line sales this moisture reduction may be desirable. If the residue gas moisture content is unimportant, the provision of cooling by an intercooler a few trays below the top of the absorber is more efficient as most of the cooling then lowers the effective absorption temperature with less loss of refrigeration to residue gas. Where very low absorber temperatures are desired some of the refrigeration duty may be applied to the incoming gas stream. MTD's in coolers are thus improved and more uniform absorber temperatures obtained. Below the 60° to 70° F. range trouble with hydrate formation in the absorber is likely. Dehydration of the incoming gases and production of a moisture free lean oil by the use of reboilers instead of direct steam stripping are effective protective measures.

High pressure absorbers are less selective between light and heavy components recovered than are low pressure absorbers. In the scheme shown on diagram 3, rich oil from a high pressure absorber is flashed to a lower pressure, and unwanted methane and ethane are to considerable degree released, thus avoiding difficulties later in condensing stripping still overhead. Desirable propanes and heavier are vaporized in appreciable amounts, however, and current practice is to provide a few trays on the top of such flash tanks to reabsorb the heavier components with a lean oil stream, as shown in diagram 4.

Another route for discarding undesired light components from the rich oil stream is shown in diagram 5. Cold rich oil is contacted in a bubble plate stripping tower with a vapor rich in propanes-butanes from vent gas sources. Ethane and lighter components are selectively stripped from the oil, being replaced with heavier components absorbed from the stripping gas stream. Released gases and remaining stripping vapors flow to a reabsorber, which is often combined with the stripper. Rich oil from the reabsorber falls into the top of the stripper. The scheme has the important advantage that it costs very little in operation.

Rich oil feed to stripping stills is preheated to vaporize absorbed light hydrocarbons. As shown in diagram 6, the preheater load may be reduced with a flash drum where vapors are separated to flow directly to the still with only the remaining rich oil flowing through the preheater. No pump is required. The reduced volume through the preheater lessens pressure drop problems, often a serious factor in plant revisions.

Two stage stripping is one effective way for avoiding expensive recompression of noncondensables. As shown on diagram 7 the first still is operated at sufficient pressure above a reabsorber so that overhead vapors not condensed do not require recompression. First still bottoms temperature is set so that second still overhead can be totally condensed with the available cooling water and at pressures low enough for effective stripping of the lean oil. First still overhead includes the bulk of butane and lighter hydrocarbons from the rich oil with possibly a little pentane and heavier. This material can be deethanized under easy conditions, especially with respect to base temperature requirements.

To extract all pentanes plus fractions from absorber gases it is essential that pentane free lean oil be employed. Gasoline plants designed for high propane extraction require lean oil circulation for propanes greatly in excess of that required for complete absorption of pentanes. In "multiple absorption," indicated on diagram 8, enough pentane free lean oil is supplied at the top of an absorber for the complete absorption of pentanes, and a larger volume of partially stripped lean oil is supplied to a point lower on the tower to effect the desired absorption of propanes-butanes. The requisite partial stripping of rich oil is accomplished by cold flashing in a series of drums with the final flash at vacuum pressure. Product separation and the stripping of a lean oil stream for pentane recovery are performed in a final stripping still.

ABSORPTION AND STRIPPING PROCESSES
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
1
2
3
4
5
6
7
8

Multi-Pressure Absorption Process

THE well head pressures, temperatures, and compositions of oil and gas as produced in the field are spread over a wide range of values. The methods of handling these materials above ground before delivery to a processing plant are diverse. Because of these factors, plants must be custom designed to suit the specific set of circumstances at hand. In addition to the variables of feed composition and supply conditions, plant design is controlled by the location of the production, the volume of reserves, the market opportunities and specifications for products and residue gases, the advisability and/or legal necessity for cycling, and the availability and cost of plant utilities and of skilled operators. Because of variations in wellhead pressures, and in producing methods, feed to a single plant is quite often received at two or more pressure levels. Gases may be rich or lean, and may or may not contain dissolved heavy oils. Mixtures of gas and distillate at high pressure are often sent directly to the processing plant. There is, too, a growing tendency toward the handling of crude oil in gasoline plants for stabilization to avoid losses of valuable vapor components.

Where several feed materials must be handled by a process plant, almost an endless number of alternate process schemes can be worked out with limits set only by common sense and the ingenuity of the designer. A frequent combination of feeds is that of a high pressure mixture of gas in combination with distillate having a boiling range from light gasoline through light gasoil. Another frequent combination is casinghead gas at vacuum pressure plus high pressure dry gas, with little material heavier than gasoline. Plant design in each case is altered to suit the conditions, the attempt always being to achieve the best overall economics.

One such combination is shown on the opposite flow diagram. Gas from low pressure crude oil separators was to be processed in conjunction with a mixture of gases and distillate from high pressure wells. Residue gas over fuel requirements was to be sold to a high pressure pipeline. High recovery of propane was desired because of strong market demand. Reserves and available production volume were adequate to justify a large and complex plant.

The distillate from field production is separated and vapors led to a "high pressure" 1050 psig. absorber. Low pressure feed gas is available at 60 psig. and since all residue gases above fuel requirements are to be delivered to a pipe line at about 1000 psig., one possible route would be to initially compress this low pressure gas to full 1050 psig. absorption pressure. The designer worked out an alternate solution, providing a second "low pressure" absorber at 400 psig. into which the compressed gases are led. Tail gas from the low pressure absorber, substantially less than originally compressed, is boosted to pipe line pressure, and worthwhile compressor horsepower savings achieved. The low pressure absorber also serves as a "reabsorber" into which rich oil from the high pressure absorber is flashed. Distillate liquid from the plant inlet separator is flashed to 400 psig. in a separate flash drum, and vapors are led to the low pressure absorber which serves as a reabsorber.

Rich oil from the low pressure absorber is in turn flashed into a reabsorber at 250 psig. Liquid from the 400 psig. distillate flash drum is flashed into a 250 psig. flash drum. Vapors from the 250 psig. flash drum are routed to the 250 psig. reabsorber.

Processing of the rich oil is not further interconnected with the processing of distillate in this particular plant, as a separate distillate stripper is provided to receive the distillate from the 250 psig. distillate flash drum.

Rich oil from the 250 psig. reabsorber is handled in a more or less conventional manner for the balance of the cycle. It is led to a stripper-reabsorber for contacting with propane rich gases to help release ethanes. The rich oil is stripped of absorbed product in two stages with the final stage operating at low pressure and with totally condensed overhead.

Light components including some pentanes are condensed overhead from the first stripping still and charged to a deethanizer. Noncondensables are recycled to the stripper-reabsorber as propane-rich vapors. Heavy gasoline is condensed as the second still overhead. Overhead vapors from the distillate stripper are routed to the reabsorber for recovery of their heavier components.

The products are further separated by a series of fractionating columns into propanes, butanes, gasoline, and a distillate fuel cut.

Residue gas flows to solid bed dehydration equipment where its moisture content is reduced to pipe line requirements.

In many such plants the distillate is not handled separately, but is combined with the rich oil. The arrangement saves in plant investment by eliminating flash drums and the distillate stripper together with preheating and condensing equipment and pumps. The combining of rich oil and distillate does present certain difficulties, however, as it is then necessary to separate the distillate product from the lean oil. Heavy distillates can not be readily vaporized for condensation as stripper still ovehead, but may require production as a side stream or bottoms product.

MULTI-PRESSURE ABSORPTION PROCESS
RESIDUE GAS COMPRESSORS
GAS TO SALES
H.P. ABSORBER
INLET GAS COMPRESSORS
LOW PRESSURE GAS
HIGH PRESSURE GAS
L.P. ABSORBER
STRIPPER-REABSORBER
REABSORBER
FUEL GAS
1ST STRIPPING STILL
2ND STRIPPING STILL
DISTILLATE STRIPPING STILL
STEAM
DEETHANIZER
DEPROPANIZER
DEBUTANIZER
BUTANES
GASOLINE
PROPANES
DISTILLATE
1 BARRELS PER DAY
2 GALLONS PER MINUTE
3 POUNDS PER HOUR
4 MOLS PER HOUR
5 STANDARD CUBIC FEET PER DAY
6 PRESSURE, POUNDS PER SQUARE INCH
7 TEMPERATURE, DEGREES FAHRENHEIT

1 of more than 200.

DURING its 18 years of continuous service to the natural gasoline and refining industries, Petroleum Engineering, Inc. has designed and constructed more than 200 gas processing plants. This cycling plant, built for Shell Oil Company at Sheridan, Texas, is typical of the various installations engineered to meet the processing problems of these industries throughout the United States and in foreign countries. The same trained technicians who designed and built this plant are available for consultation on your processing problems.

PETROLEUM ENGINEERING, INC.

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Photographed for Petroleum Engineering, Inc. by Robert Yarnall Richie

Absorption Oil Purifying Process

PETROLEUM ENGINEERING, INC.

THE accompanying flow diagram shows an Oil Purifier Unit used in absorption plants, in the petroleum and coke-oven-gas industries. In the latter industry, absorption plants are used for extracting liquid aromatics from coke-oven gases.

This unit continuously removes suspended solids, waxes, gums and the higher mol-weight undesirable hydrocarbons from the circulating stream of absorption oil. Undesirable solids and heavy liquids accumulate continuously in the purifier unit and are periodically withdrawn from the system and discarded. The undesirable material is removed from the purifier unit, while hot, in form of heavy sludge, sometimes so viscuous it could not be removed unless hot. Ordinarily it has the consistency of road oil and frequently is used for that purpose around the plants. In coke-oven-gas plants withdrawal is frequent as compared to plants in the oil industry due to the nature of the gas processed. In natural-gasoline absorption plants, withdrawal may be every two weeks or possibly once a month.

In most plants it is possible to keep the absorption medium practically water white. This results in clean heat-transfer surface, oil coolers, heat exchangers and preheaters. Clean heat transfer apparatus means higher still and lower absorber temperatures, and improved efficiency.

Since heavier hydrocarbons are removed from the absorption oil, the result is a lower molecular weight, and therefore more efficient absorbent. Such an oil is easier to strip, so less process steam may be used or better stripping obtained with the same steam.

The purifier unit is operated in conjunction with the conventional still, may be considered a simple still, but it does not incorporate a condenser. The stripping section of the regular absorption-plant still serves as the condenser. Distillation is carried on at a temperature and pressure approximating the base temperature and pressure of the absorption still with which it works. The temperature is maintained by use of an internal heating bundle heated by steam or hot oil. All or the greater portion of the open process steam which is to be used in the plant still is passed through the purifier unit. Since the quantity of material evaporated in the purifier unit is relatively small as compared to the amount of material evaporated in the main still, the quantity of process steam passed through this unit is proportionately large. This means that absorption oil can be distilled under temperature and pressure conditions that would not be economical for separate distillation. Vapors from the purifier unit pass through a mist extractor and into the normal stripping steam connection of the absorption-plant still. These vapors are principally steam, but include a substantial quantity of oil vapors. The steam performs the usual function in the absorption-plant still and the oil vapors are recondensed on the stripping trays. In being recondensed these oil vapors give up heat at a point in the main still where the addition of heat is useful.

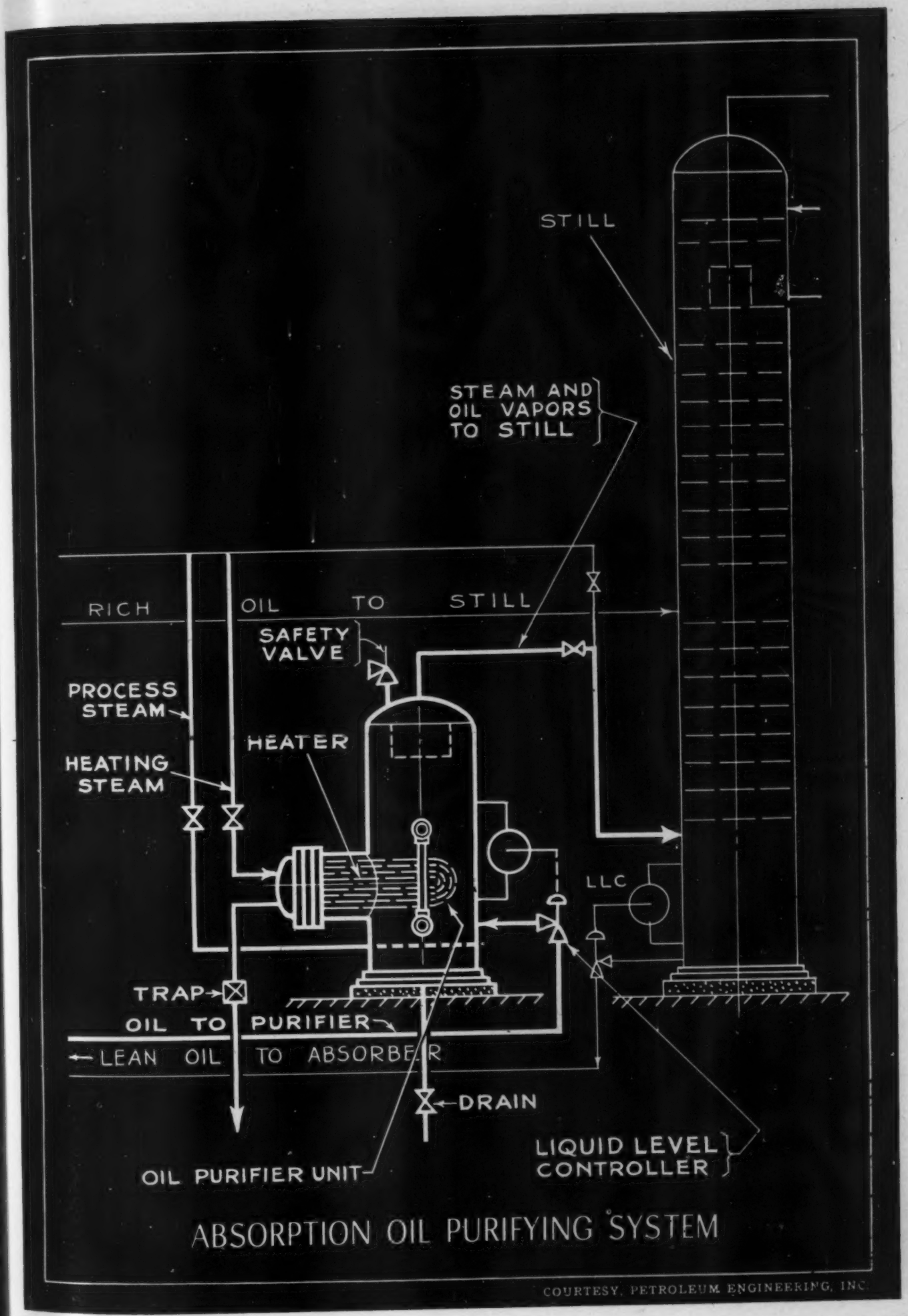
The quantity of oil distilled in this unit depends upon temperature and pressure, the quantity of the steam passed through the unit, and the physical characteristics of the oil remaining in the unit. Distillation rate gradually tapers off during the period of concentration of heavier materials in the purifier and it is the usual practice in most plants to dump the purifier unit when the distillation rate has been reduced to around 0.5 or 1.0 percent of the total plant circulation. Distillation rate can be checked by passing a sample of the purifier vapors through a condensing coil and comparing the quantity of water with oil condensed.

The only automatic control is a liquid-level controller, which maintains a level in the unit by admitting oil from the plant system as fast as oil is evaporated and released. No control is applied to the heating medium. If steam is used for heating, the full line pressure is allowed on the bundle, and the rate of distillation is not limited. In most plants, distillation rate is considerably higher than necessary for maintenance of absorption-oil condition. There is no reason however, for limiting the performance of the unit since its operation is advantageous to the operation of the main still in that heat is added to the stripping section.

No separate process steam is used. Whether steam or hot oil is used, the heat is recovered in the stripping section of the still. It may therefore be considered that the only extra plant utility required is heat lost in radiation from insulated surface. Heat radiation from the unit proper and hot piping that otherwise would not be installed in the plant, is negligible.

This oil purifier (patented) is designed and built by Petroleum Engineering, Inc., of Houston and Tulsa.

Since the initial tryout of this unit, this purifier has been incorporated as standard in absorption plants constructed by Petroleum Engineering, Inc., and several other plant designers. It also has been added to plants built before its inception in 1940.



ABSORPTION OIL PURIFYING SYSTEM

COURTESY, PETROLEUM ENGINEERING, INC.

Gasoline and Gas Fractionation Process

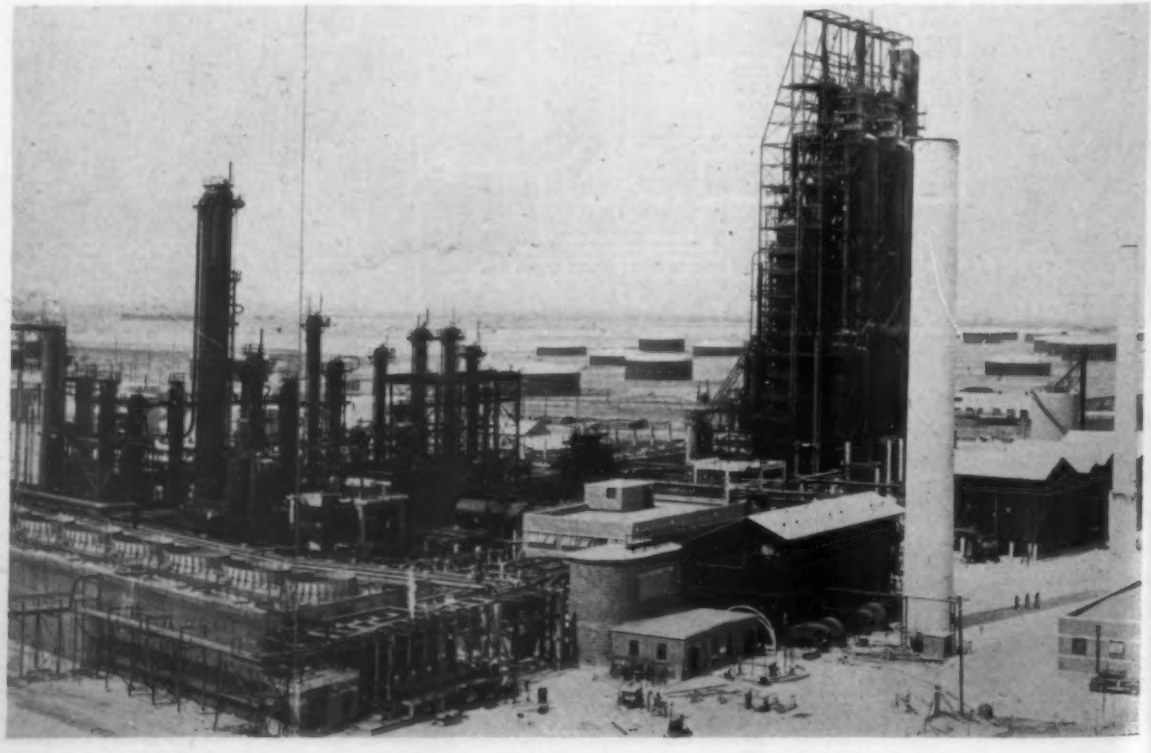
THE LUMMUS COMPANY

GAS-CONCENTRATION and stabilization plants are among the earliest secondary processes used in petroleum refining. The necessity for such plants dates back to the earliest days of thermal cracking, especially in situations where low-pressure separation of the cracking-plant effluent was conducted. As the art of thermal cracking developed there was a trend towards employing high-pressure-fractionation systems and thereby minimize the necessity for gas-recovery systems.

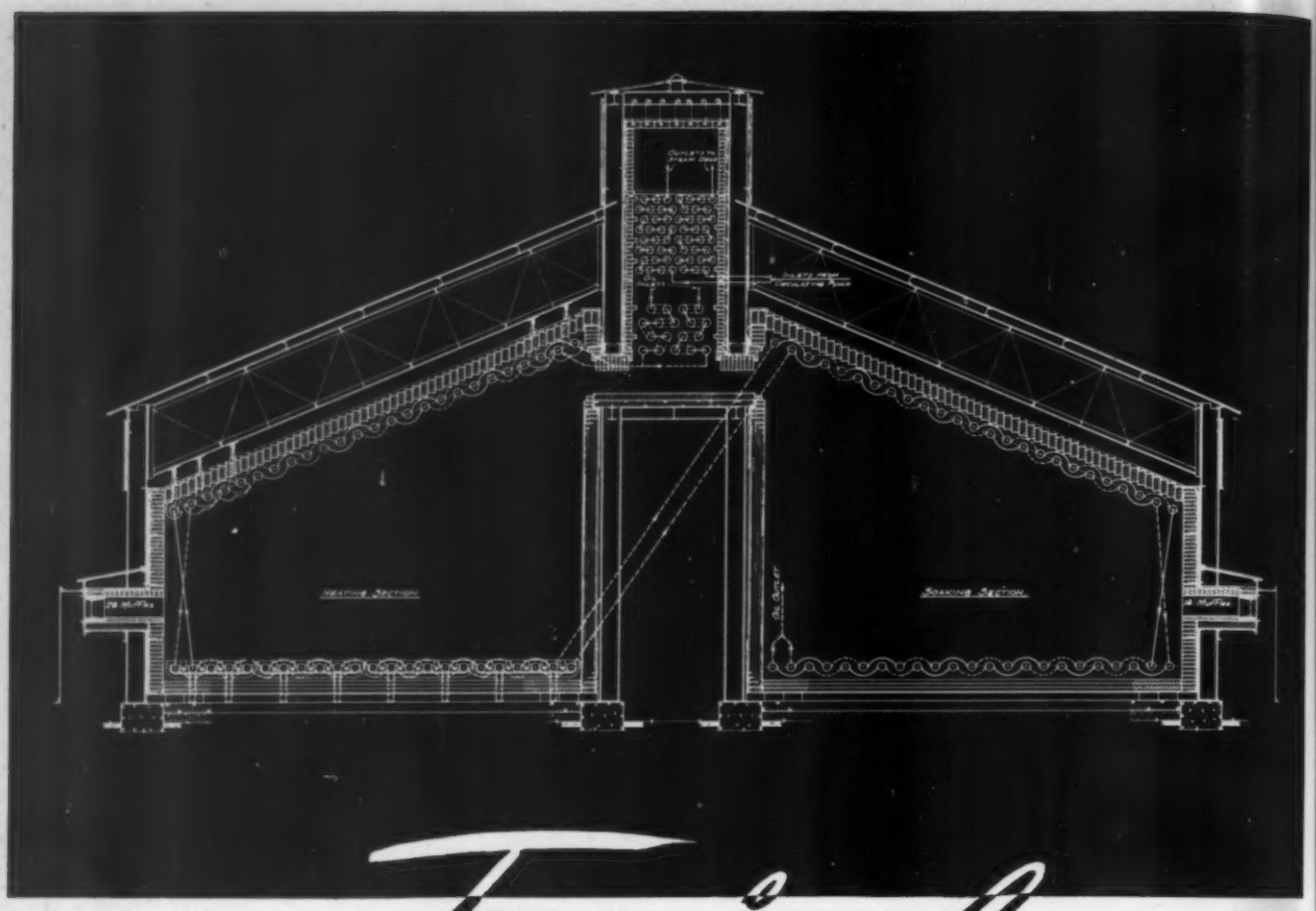
Practically all refineries in operation today include some form of gas-recovery and stabilization plant. The design of this plant depends to a considerable extent upon the type of processing units employed and the amounts of wet gas and unstable distillates to be handled. However, there have been significant advances in methods of handling these materials, with the result that substantial savings in both investment and operating costs can be made over antiquated or obsolete systems. Each refinery situation requires special consideration.

The advent of catalytic cracking brought about a great increase in the installation of gas-concentration and stabilization facilities for the twofold reason that catalytic-cracking plants produce large quantities of pentane and lighter materials and that these plants must of necessity fractionate the reactor effluent at low pressure. The most suitable type of gas-concentration facilities for operation in conjunction with catalytic cracking will depend to some extent upon the type of catalytic-cracking operation and the nature or the charging stock. Gas-concentration units of a variety of designs, however, may be used for accomplishing any required separation and recovery. The Lummus Company has given careful consideration to the many possible types of plant and has determined that the system shown in the flow sheet opposite represents one of the simplest and most economical for situations where it is necessary to recover all C_4's and high recovery of C_3's.

This plant consists of a compression system for raising the pressure of the wet gas to approximately 150 psig and contacting this material with the unstable distillate; an absorber for processing the wet gas from the contact drum with debutanized gasoline and heavy naphtha; and a debutanizer for producing debutanized pasoline and an overhead C_3, C_4 condensate stream. The debutanizer in effect acts as a still for the absorption naphtha, however part of the lean oil flows directly from the low-pressure-fractionating system of the catalytic-cracking unit to the top of the absorber for the purpose of sponging the small amount of pentanes and heavier material evaporated from the debutanizer gasoline making up the bulk of the lean oil. The debutanizer overhead condensate may be charged through a Girbotol unit to a catalytic polymerization plant or may be charged to a depropanized for recovery of butane-butylenes and an overhead C_3 condensate. The debutanizer may be reboiled with 450-pound steam or with hot circulating oil from the low-pressure-fractionating system, the latter usually being more desirable. In some situations it has been found desirable to provide a fractionating absorber equipped with a reboiler in order to reduce the gas recycle from the debutanizer and the amount of gasoline recycled to the absorber for lean oil.



LUMMUS GAS CONCENTRATION UNIT
COMPRESSOR
ABSORBER
DEBUTANIZER
DEPROPANIZER
WET GAS & UNSTABLE GASOLINE FROM CATALYTIC CRACKING UNIT
FUEL GAS
HEAVY NAPHTHA FROM CATALYTIC CRACKING UNIT
DEBUTANIZED GASOLINE
450# STEAM
150# STEAM
GIRBOTOL PLANT
BUTANES
CONDENSATE
OFF GAS
COURTESY, THE LUMMUS COMPANY.

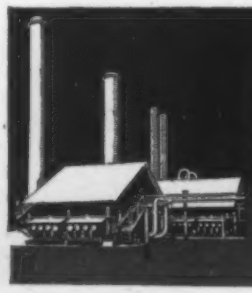


Typical

ALCORN DOUBLE PURPOSE HEATER

This tailor-made heater with controllable heat curve for thermal cracking or reforming is a typical Alcorn installation, one of many types which Alcorn makes to meet every requirement for petroleum processing.

Alcorn's 24 years experience is available for the solution of your heating problems. Send your specifications for new installations or replacements.



ALCORN
Combustion Company
SCHAFF BUILDING PHILADELPHIA
Los Angeles • Houston • San Francisco • Tulsa

TOMOCONNOR

Setting new standards in gasoline plant efficiency

Ground broken, February 1947 — compressors operating, October—Entire plant on stream, December 1947. Designed to process 35,000,000 cubic feet of gas daily at pressures ranging from 65 lbs. to 850 lbs. and employing 7500 horsepower.

Tomoconnor adds another important chapter to the long record established by Gasoline Plant Construction Corporation for outstanding achievement in modern gasoline plant engineering.

A competent engineering staff composed of specialists in Petroleum, Chemical, Mechanical, Electrical and Civil Engineering is maintained to provide specialized and thoroughly coordinated service in engineering, design and construction of complete plants or components to fit your individual requirements.

FOR MAXIMUM EFFICIENCY IN NATURAL GAS PROCESSING LET US PLAN YOUR NEXT PLANT.

GASOLINE PLANT CONSTRUCTION CORP.

ENGINEERS DESIGNERS CONSTRUCTORS

Vapor Recovery Process

FOSTER WHEELER CORPORATION

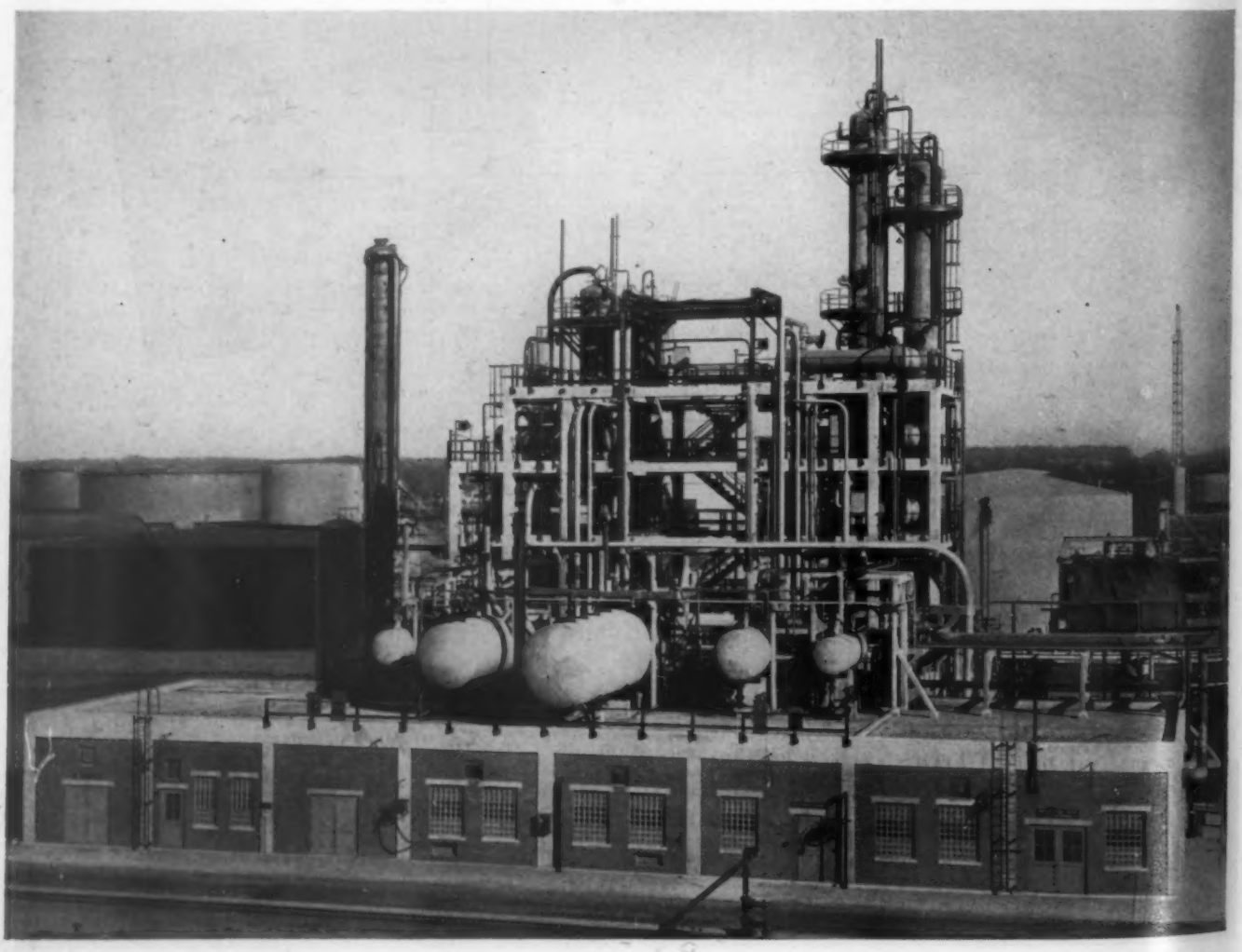
THE three refinery operations, absorption, debutanization-stabilization and rerunning were, until recent years, three separate process units and were generally located in different parts of the refinery. However, the modern trend is to combine these processes in one self-contained unit, thereby affecting substantial reduction in labor and utility costs and initial capital expenditure.

In the diagrammed typical vapor-recovery unit, employed very successfully to process fluid catalytic cracking unit products, the gas and liquid charge is united in the charge tank. C_2's and lighter are removed from the system as gases from the top of the absorber with the heavier gaseous products being absorbed by the lean oil and returned to the charge tank. The total liquid is then charged to a high-pressure still where the naphtha is partially stabilized and charged to the low-pressure still for complete stabilization. The net overhead product from both the high- and low-pressure stills are joined in the high-pressure still accumulator and constitute the charge for the rectifier and debutanizer.

The high- and low-pressure still system is installed in place of a single still in order to avoid the use of high-pressure (600 psig) steam for reboiling.

Under this system the total overhead stream is condensed at 100 psig and the stabilized bottoms product reboiled at 30 psig. This permits the use of 100-psig steam in the reboilers and normal plant cooling water in the condensers. The bottoms product from the low pressure still is a 2-pound R.V.P. gasoline, part of which is used as lean oil in the absorber and recycled to that tower after being cooled to 90° F.

The net overhead product from the still system is charged to the rectifier from which propane is taken as the overhead product. The bottoms from this tower is charged to the debutanizer, the last tower in the vapor recovery system. The overhead product, a butane: butene cut, is a valuable stock for polymerization and alkylation. The bottoms product is a debutanized light gasoline fraction.



VAPOR RECOVERY PROCESS
REFINERY FUEL
PROPANE
BUTANE
DEBUTANIZED ABSORPTION NAPHTHA
STABILIZED HEAVY NAPHTHA
DRY GAS
LEAN OIL
CHARGE
CONDENSER
REBOILER
COOLER
HEATER
HEAT EXCHANGER
ABSORBER
REFLUX
PMD-888-A
AD-1468
COURTESY, FOSTER WHEELER CORPORATION

Gasoline and Gas Fractionation Process

E. B. BADGER & SONS COMPANY

DEVELOPMENTS in fractionation and light-ends processing from cracking-plant products received an impetus during the war from the demands for stabilization of aviation base stocks and segregation of light ends for alkylation or other subsequent processing.

Outlined in the flow diagram is a gas plant designed and constructed by E. B. Badger & Sons Company for a large refiner in this country. This plant is designed to receive gas and gasoline from a catalytic cracking unit and from a catalytic treating unit, and to produce 30,000 barrels per day of combined liquid products as well as 6,500,000 cubic feet of fuel gas. Debutanized motor gasoline, specification aviation gasoline, commercial propane, butanes-butenes for alkylation or poly-unit feed, and isopentane are the principal end products from the unit.

Motor gasoline from the cracking plant is charged to the debutanizer. This column is operated at 185 psig to effect recovery of the major portion of the overhead as liquid, and the debutanized gasoline goes to storage.

Aviation gasoline from the treating unit is stabilized at approximately the same pressure, also to obtain maximum recovery of the overhead as liquid, and the stabilized aviation is pumped to storage. The liquid overhead products from the debutanizer and stabilizer are pumped to a surge tank from which they are sent to the de-ethanizer.

Cracking-unit gas is transmitted to this plant at 50 psig. The smaller amount of treating-unit gas, available at lower pressure, is compressed to 50 psig, and these streams are combined with the small flow of overhead gas from the debutanizer and stabilizer, compressed in two stages and discharged directly into the de-ethanizer which operates at 450 psig. The overhead from this column at 125° F. is passed through a water-cooled condenser and then is refrigerated, providing reflux at a temperature of 45 to 50° F., the possibility of hydrate formation precluding the use of lower temperatures. A feature of the refrigeration unit design is the use of propane produced in the plant as refrigerant. The refrigeration of the de-ethanized overhead condensate is effected by the evaporation of plant propane in a kettle-type tubular unit, the propane vapor so produced being compressed and then condensed in the depropanizer overhead product condenser. Net overhead product from the de-ethanizer passes to the refinery fuel-gas system as a vapor.

De-ethanizer bottoms are delivered to the depropanizer which operates at 250 psig. In this column, commercial propane is produced overhead. The bottoms are sufficiently depropanized so that the subsequent butane product will contain less than 1 mol percent propane. The butane tower operates at 100 psig and is designed to make a sharp separation between normal butane in the overhead and isopentane in the bottoms. The bottoms from the butane tower is feed for the isopentane column. This column, operating at 22 psig, is designed to produce overhead an isopentane fraction of 98 percent purity. The bottoms from the isopentane column comprising normal pentane and heavier are pumped to storage and used in motor-gasoline blends. Although two isopentane towers are shown in the flow diagram, it will be noted that process-wise they actually constitute one tower. A split tower was used since the height of a single tower would have been excessive for the particular installation here discussed.

LIGHT ENDS FRACTIONATION AND STABILIZATION
COURTESY, E. B. BADGER & SONS COMPANY
DEBUTANIZER
STABILIZER
DEPROPANIZER
BUTANE TOWER
ISO PENTANE TOWER
TO FUEL
STABILIZED AVIATION GASOLINE
DEBUTANIZED MOTOR GASOLINE
PROPANE PRODUCT
BUTANE PRODUCT
PENTANE + HEAVIER
ISO PENTANE PRODUCT
CAUSTIC
TREATING ACID

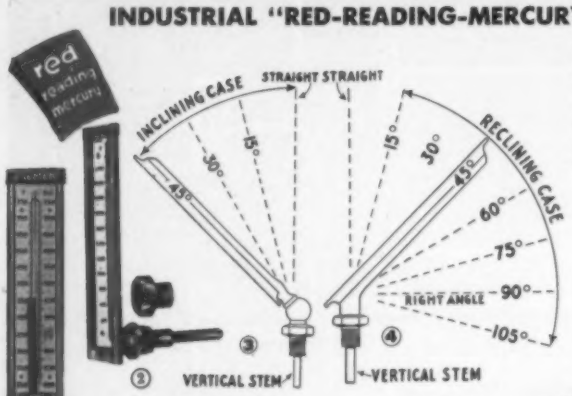
PALMER THERMOMETERS, Inc.

Manufacturers: Industrial, Laboratory, Recording and Dial Thermometers

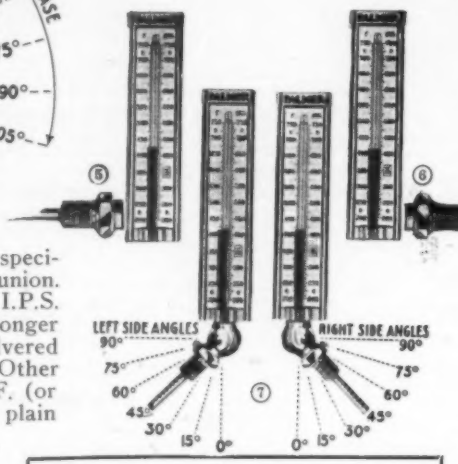
Main Office and Factory:
2513 Norwood Ave., Cincinnati (Norwood) 12, Ohio

Canadian Branch Factory:
King and George Sts., Toronto 2

INDUSTRIAL "RED-READING-MERCURY" THERMOMETERS WITH NEW EXTRUDED BRASS CASE



PALMER Industrial Thermometers feature a new revolutionary Palmer Patent —an extruded brass case in 4", 7", 9", and 12" sizes. Provides larger reading scale and longer service. Each tube is made from selected glass, with "Red-Reading-Mercury" feature, a pure mercury tube with a wide RED column, easier to read. Accuracy guaranteed.

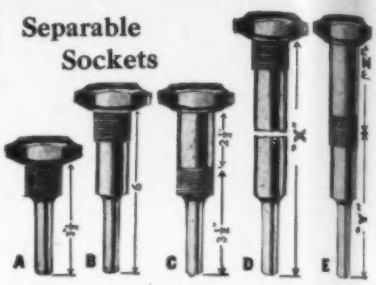


Dimensions: **Case** . . . 7", 9", 12". **Fittings** . . . Fixed taper thread, Union, Separable-socket (usually specified). **Flanges:** fixed, adjustable or union. **Thread Size** . . . ¾" and 1" I.P.S. **Stem** . . . 3½" including thread and longer lengths. **Finish** . . . White or silvered scale; Nickle-plated or chrome mounting. Other finishes when specified. **Ranges:** —40 + 750° F. (or C.). Up to 1000°. with borosilicate glass and plain mercury tube. Illustrations above:

1. Straight (with socket)
2. Angle (with Union)
3. Diagram: Inclining case with Vertical stem.
4. Diagram: Reclining case with Vertical stem (also used as Inclining case with horizontal stem).
5. Left Side Angle.
6. Right Side Angle.
7. Diagram: Oblique stem with Side Angle case.

REPAIRS . . . To all makes of mercury Industrial Thermometers, furnishing "Red-Reading-Mercury" tube (no extra cost) and replacing all worn or broken parts, making Thermometer as good as new. Guaranteed Accuracy. New extruded case included.

Separable Sockets



Palmer sockets are interchangeable with sockets of similar length. Can be furnished in various metals, according to the solution to come in contact with socket. Standard metals: Brass, Steel, Stainless-steel.

A. Standard length 3½" under hex.
B. Measures 6" under hexagon.
C. With 2½" extension neck, total 6" under hex.
D. Misc. length socket. Give "X" dimension.
E. Misc. length socket with extension neck. Give "Y" and "Z" dimension, when ordering.

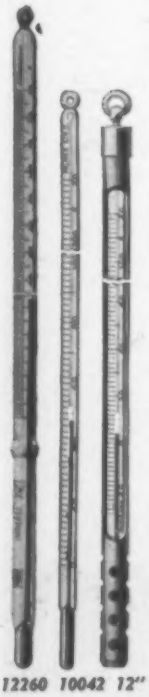
POCKET THERMOMETERS

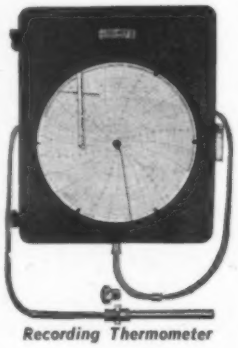
. . . with "Red-Reading-Mercury" feature. Handy to carry around for quick tests. Round or Lens tube. 6" overall metal case with pocket clip. Guaranteed accurate.

13240

LABORATORY THERMOMETERS

—Engraved glass thermometer tube with "Red-Reading-Mercury." The wide RED column is much easier to read. Round or lens glass; Ring top; Button top for armours; Total immersion or standard 3" Partial Immersion; Other depths special; Graduations: 1°, 2°, 5° or Fractional Division. Each packed in separate box. Guaranteed accurate.


12260 10042 12"

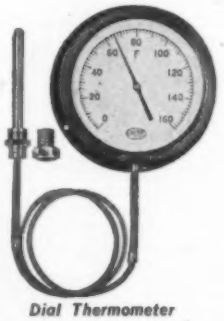

Recording Thermometer

"SUPERIOR" RECORDING THERMOMETER

Mercury Actuated. Very sensitive and Guaranteed Accurate. Every part constructed for long and satisfactory service. 12" die-cast aluminum case; black or white wrinkle or satin finish; all parts rust-proof; stainless-steel bulb and flexible armoured tubing up to 100 ft. Fittings: plain, Union, separable-socket (usually specified), flanges: adjustable or union. Electric or Spring clock. 12 or 24 hr. or 7-day charts, easy to remove. Ranges: —40 + 1000° F. (or C.) Standard charts for all types of ranges.

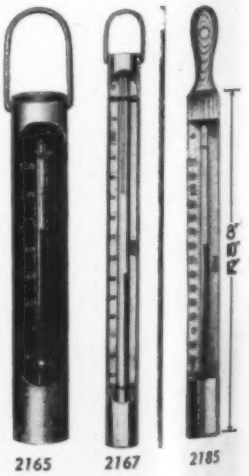
METAL ARMOURS . . .

As an extra protection for Laboratory Thermometers, when in use, metal armours can be furnished (as illustrated). ⅜" O.D. Perforated end. Screw top with ring. Nickeled finish. Ranges up to 750° F. Brass armour standard. For high ranges, steel armour furnished. If stainless-steel is wanted, specify on order.


Dial Thermometer

"SUPERIOR" DIAL THERMOMETER

Mercury Actuated. Guaranteed Accurate. Very Sensitive. Built sturdy, to give long service. 8" round case: rubberized black finish; all parts rust-proof; Bulb and flexible armoured tubing up to 100 ft. made of stainless-steel; Fittings: Plain, Union, Separable-socket and adjustable or Union-flange; Range: —40 + 1000° F. Standard dials for all ranges.



2165 2167 2185

HYGROMETERS & PSYCHROMETERS:

For testing Humidity, we furnish Wet & Dry Bulb wall style Hygrometer or the Sling style Psychrometer. Standard Ranges: + 30 + 120° F. and + 80 + 220° F. With "Red-Reading-Mercury" tube, so easy to read.

CUP AND ROOM THERMOMETERS:

With lens tube, "Red-Reading-Mercury" feature. Graduated Scale. Guaranteed Accurate. Easy to Read.

A.S.T.M. and PRECISION ALL-GLASS THERMOMETERS,

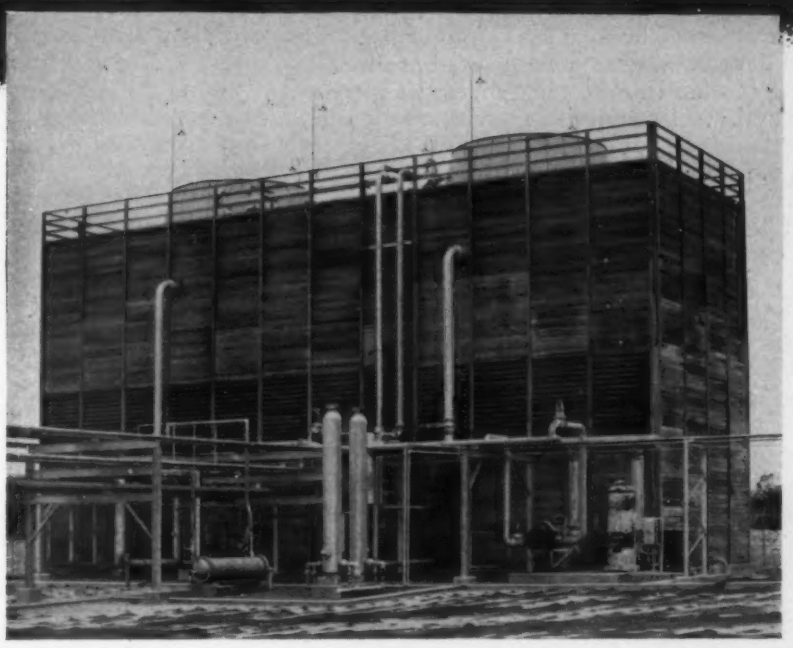
with "Red-Reading-Mercury." Made to definite specifications and acceptable to U. S. Bureau Standards.

For Good Thermometers—Specify Palmer

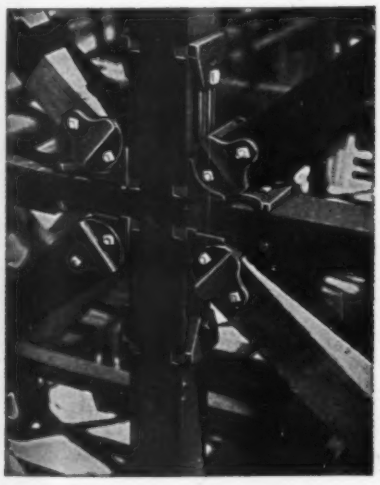
Efficient Tower Design...

PRODUCES "COOLER" PROFITS

Proper design and rigid time-resistant construction assures colder water during a longer service life. This all adds up to lower operational costs—which means "cooler" profits from your processing operations.

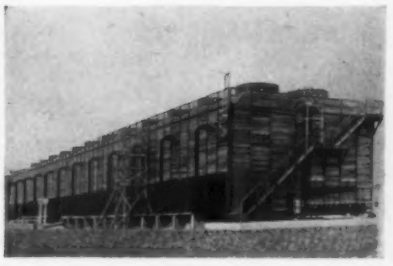


Two cell, coil shed type, redwood induced draft cooling tower serving recycling plant for major oil company.



TIMBER CONNECTORS

Patent No. 2,280,121—Special cast iron timber joint connector used to develop the full strength of the timber joint, an accomplishment almost impossible to achieve by bolting only.



Double induced draft cooling tower equipped with twenty-four 12 foot diameter adjustable pitch propeller type fans.

Standard Redwood Induced Draft cooling tower, equipped with adjustable pitch, propeller type fans driven through our specially designed right angle spiral bevel gear reduction units by motors located outside the air stream.

For Induced Draft, Forced Draft, Atmospheric Deck or Spray Towers of design and construction based on sound engineering principles . . . to meet specific requirements for cooling performance and structural strength, consult

WATER COOLING EQUIPMENT COMPANY

8605 New Hampshire Avenue **Affton Station, St. Louis 23, Mo.**

Naphtha Fractionation Process

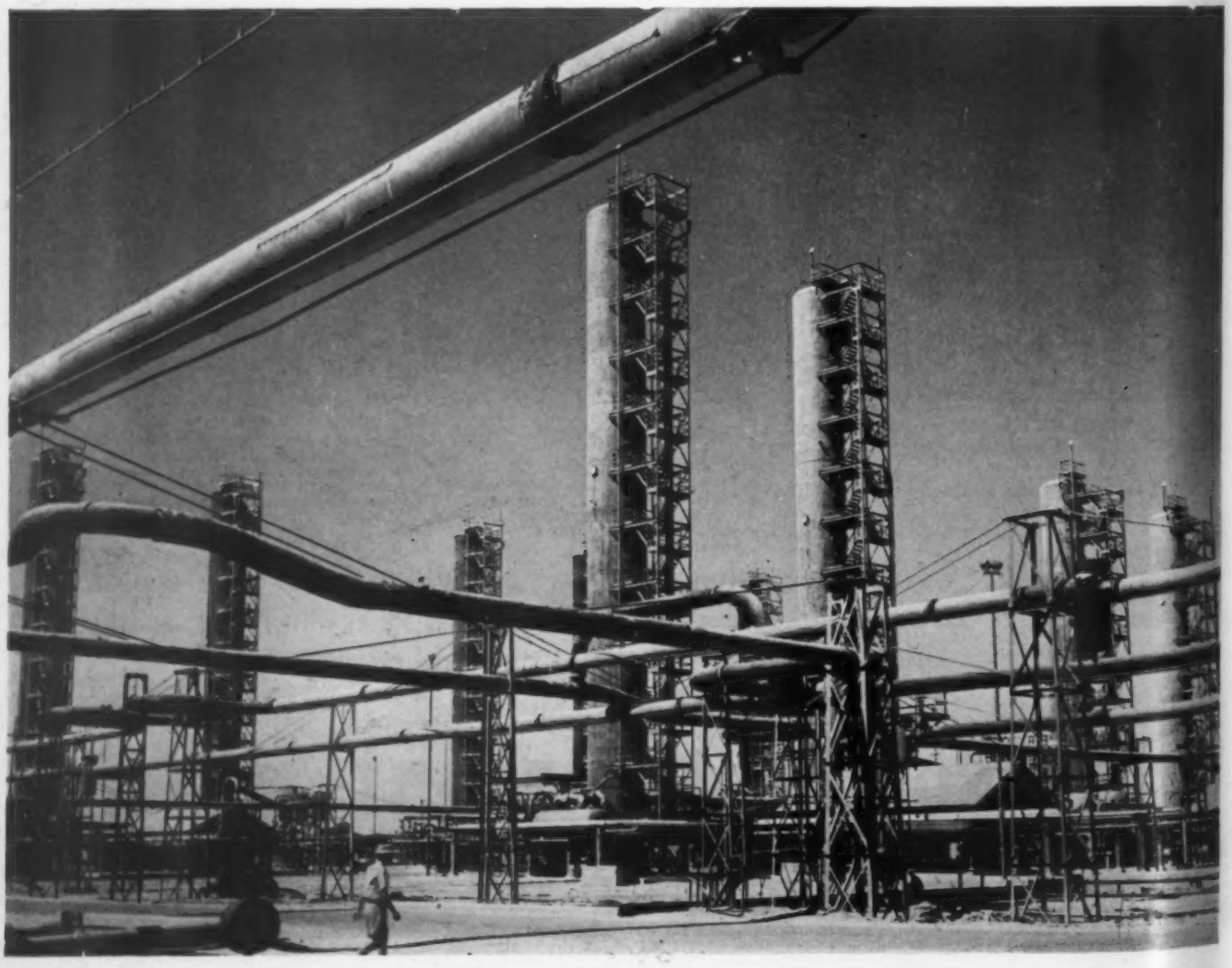
E. B. BADGER & SONS COMPANY

SUPERFRACTIONATION of naphtha fractions to separate the high-octane branched-chain components from the low-octane components was one method developed during the war to produce aviation blending stocks. In many instances the investment in money and steel required for aviation production by this method was but a small fraction of that required for the manufacture of high-octane materials by other means. This manufacturing procedure also may be applied to the peacetime production of high-octane motor gasolines.

In one installation by E. B. Badger & Sons Company, as illustrated by the accompanying flow diagram, a naphtha cut containing compounds ranging from normal pentane through normal heptane was charged to a series of superfractionators. Normal pentane was taken overhead in the first column and the bottoms containing isohexane through normal heptane was charged to the second superfractionator. Here isohexane was obtained as an overhead product, and the bottoms, charged to a third column, were split into normal hexane overhead, and isoheptane and heavier as bottoms. The isoheptane was taken as the overhead in the last column.

In some cases, due to the large number of plates and the resultant height, a fractionating column was split into two sections, with the vapor from the first section passing into the base of the second section, and with reflux pumped from the base of the second section into the top of the first section. In the flow sheet, for simplicity, the columns are shown as single units.

Accurate control of the distillation conditions is essential to the success of fractionation operations of this type, and Badger engineers developed the special instrumentation required. The use of surge tanks of ample capacity between the different columns contributes to steady flow and consistent operation when the composition of the feed varies. Due to the close boiling points of the fractions to be separated, product drawoff quantities generally are regulated by volume control, based on laboratory analyses of the feed stocks.



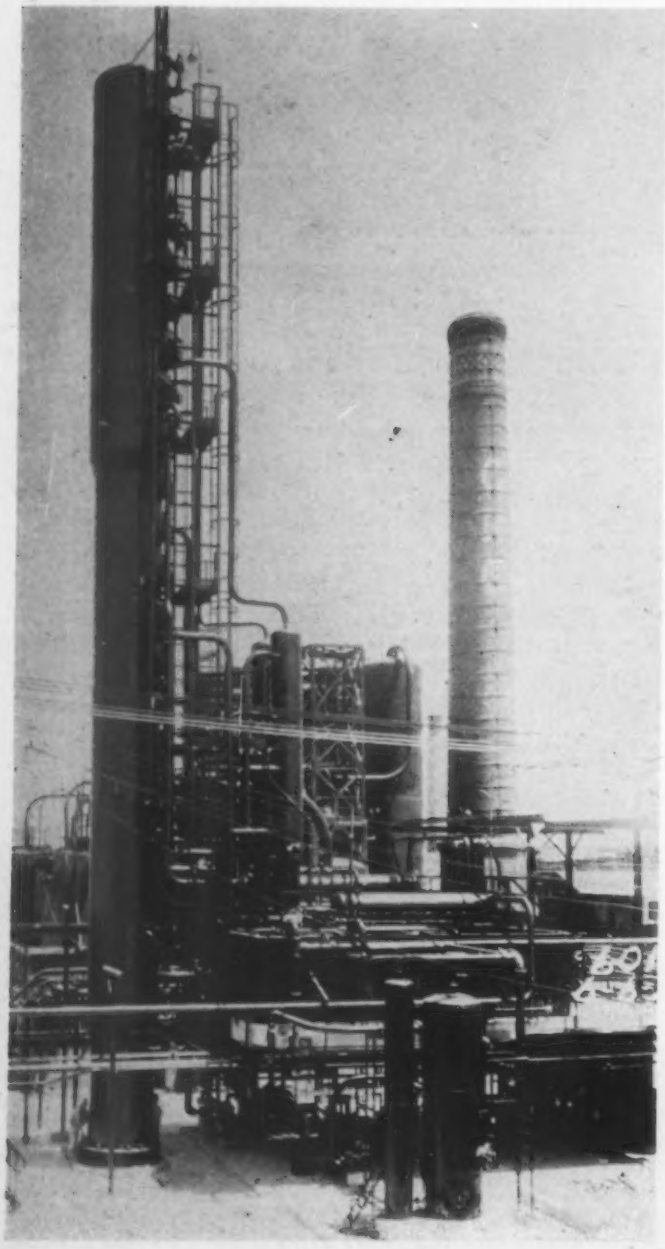
BADGER SUPERFRACTIONATION UNIT
COURTESY, E. B. BADGER & SONS COMPANY
CHARGE
PREHEATER
CONDENSER
REBOILER
NORMAL PENTANE
ISO HEXANE
NORMAL HEXANE
ISO HEPTANE
NORMAL HEPTANE

Naphtha Special Rerun Process

THE LUMMUS COMPANY

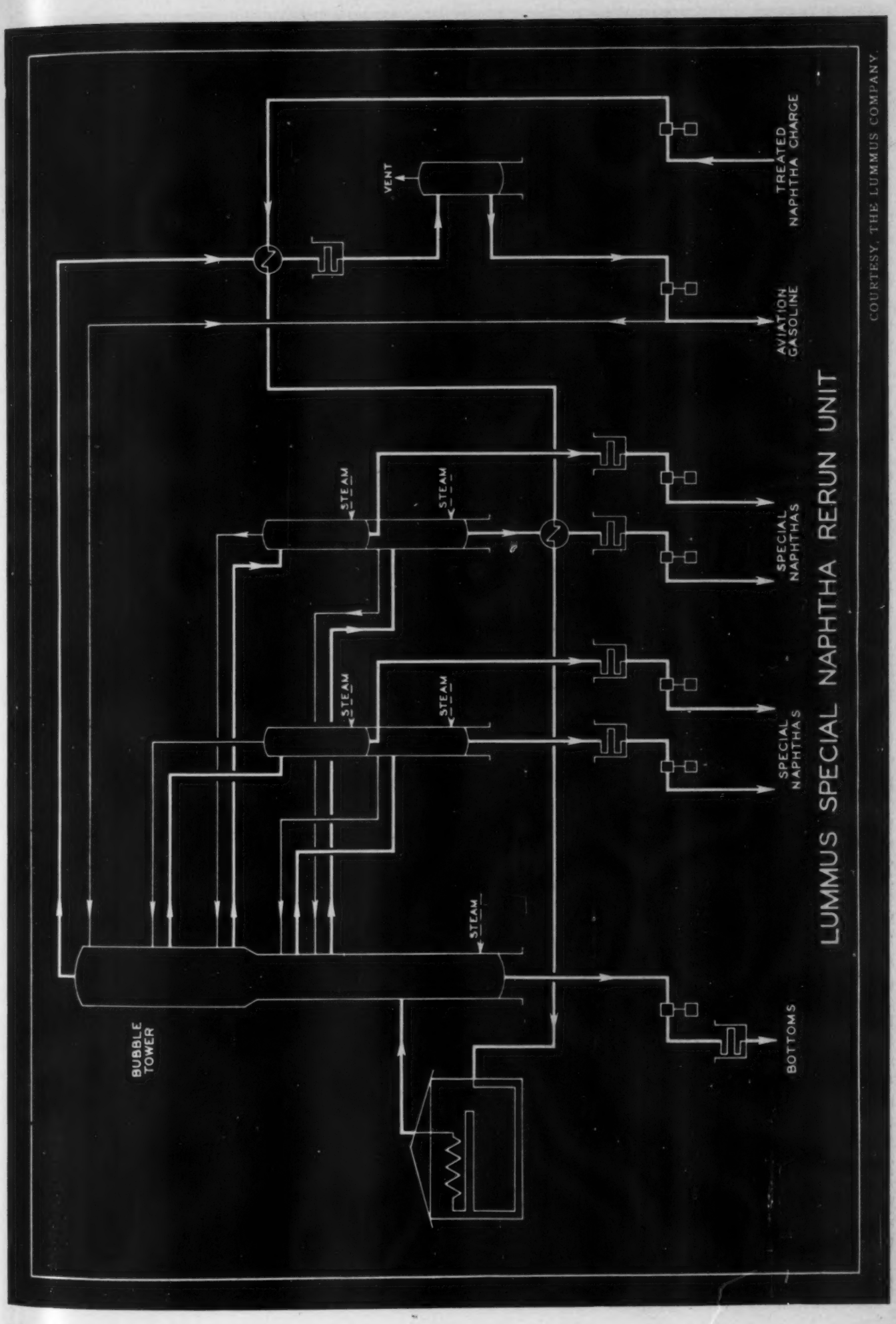
THE utility of a Lummus special naphtha-rerun unit built for a large refiner was so clearly demonstrated that the same refiner placed an order for an exact duplicate to meet his additional requirements for this service.

The engineering problem presented involved consideration of a number of factors. The chief characteristic desired was flexibility, since the unit was required to operate on a variety of naphthas and somewhat heavier distillates for production of specialty products of precise specifications. The second consideration calling for special design was introduced by the fact that material being charged to the unit had been treated and the fractions produced were to be finished products ready for marketing. For this reason it was necessary to insure that no overheating of the charge material would take place and result in off-test products. To meet these requirements a design was worked out in accordance with the flow diagram shown on the page opposite. The fractionating tower was set up for the production of an overhead stream, four side streams and a residue. The overhead product was rubber solvent or low-endpoint gasoline, depending on the characteristics of the charging stock. The four side streams were naphthas of various characteristics, most of them of narrow boiling range and well fractionated. The original intention was to produce as finished products the first and third side-stream fractions only, since the charging stock indicated that simultaneous production of more than two side streams would not enable specifications to be met. However, the performance of the unit was so much better than expectations that it was possible to produce three simultaneous side streams to specifications instead of two.



Avoidance of overheating was obtained by careful attention to the oil-heating unit. The Lummus oil heater lends itself to satisfactory solution of this problem because of its symmetry and because of the distribution of heat absorption. In view of the fact that complete vaporization of most of the various charging stocks takes place at a temperature below 450° F., it was logical to reverse normal flow of material through the heater by passing it first through the high-heat input section of radiant tubes, then through the semi-radiant roof tubes and finally through the convection bank where vaporization was allowed to take place. By the use of this flow the temperature distribution was ideal and it was possible to effect complete vaporization of the charging material without reaching a peak temperature which would be injurious to the material being handled. Further, the maximum temperature required for complete vaporization is so low that there is no sacrifice of furnace efficiency when using the reversed flow described.

Another feature of this unit which may be mentioned involves the grouping of control and metering instruments so that operation of the unit can be carried out with very little supervision and effort. To accomplish this, controls were conveniently grouped in the control room. Provision for visual inspection and flow control of all streams with remote control valves on stripping steam to the side stream strippers was provided. This arrangement was made so compact that essentially "arm-chair operation" is possible. In view of the many different services which this unit is called on to perform, the importance of close control is evident, and its value has been clearly demonstrated by the successful operation of the unit.


LUMMUS SPECIAL NAPHTHA RERUN UNIT
COURTESY, THE LUMMUS COMPANY
VENT
TREATED NAPHTHA CHARGE
AVIATION GASOLINE
STEAM
STEAM
SPECIAL NAPHTHAS
STEAM
STEAM
SPECIAL NAPHTHAS
STEAM
BUBBLE TOWER
BOTTOMS

Hypersorption Process

UNION OIL COMPANY OF CALIFORNIA

HYPERSORPTION is a process for the separation of hydrogen, light hydrocarbons and other gaseous compounds which are not readily or economically handled by conventional fractionation procedures. The process employs a moving bed of activated carbon granules and depends on the ability of this material selectively to adsorb heavier hydrocarbons or larger molecules to a greater degree than light hydrocarbons or smaller molecules. Several commercial units employing the process are now in operation with others in various stages of design and construction.

Referring to the flow diagram, it will be noted that the Hypersorption unit, termed a Hypersorber, has three major sections: a cooling section at the top, an adsorption and rectifying section in the middle, and a steam stripping section at the bottom. A gas lift is employed to convey the granular charcoal from the bottom of the unit to the top in a closed cycle of circulation.

The feed gas enters through a distribution plate and passes upward counter-current to the moving bed of activated charcoal. In its passage through the bed, the heavier compounds are adsorbed and retained on the charcoal. The top product gas, comprising the lightest constituents in the feed, disengages from the charcoal below the cooler and passes off through a cyclone. The charcoal passes downward below the feed point to the rectifying section where it contacts a reflux of heavy constituents, separated in the stripper which serves to liberate lighter constituents adsorbed on the downward-moving charcoal. A concentration of these latter constituents in the vapor phase results and if desired they can be disengaged as a side-cut product gas at a point between the feed and bottom product gas disengaging sections. As the charcoal flows past the bottom product gas disengaging section into the steaming section, it meets an upward flow of steam which liberates the adsorbed heavy components, and these flow upward either to return as reflux or pass out as bottom product gas from the Hypersorber.

The charcoal flows from the steaming section into the stripping section wherein it passes down through tubes which are heated by a circulating heat-transfer medium and through which steam is passed counter-current to the charcoal. The hottest point in the stripper is at the bottom of this section and substantially all residual adsorbed material is removed during the passage of the charcoal through this zone.

The flow of charcoal leaving the stripper is controlled by means of a specially designed reciprocating feeding mechanism which drops the charcoal into a sealing leg in which a level is maintained to restrict the flow of steam into the charcoal lift.

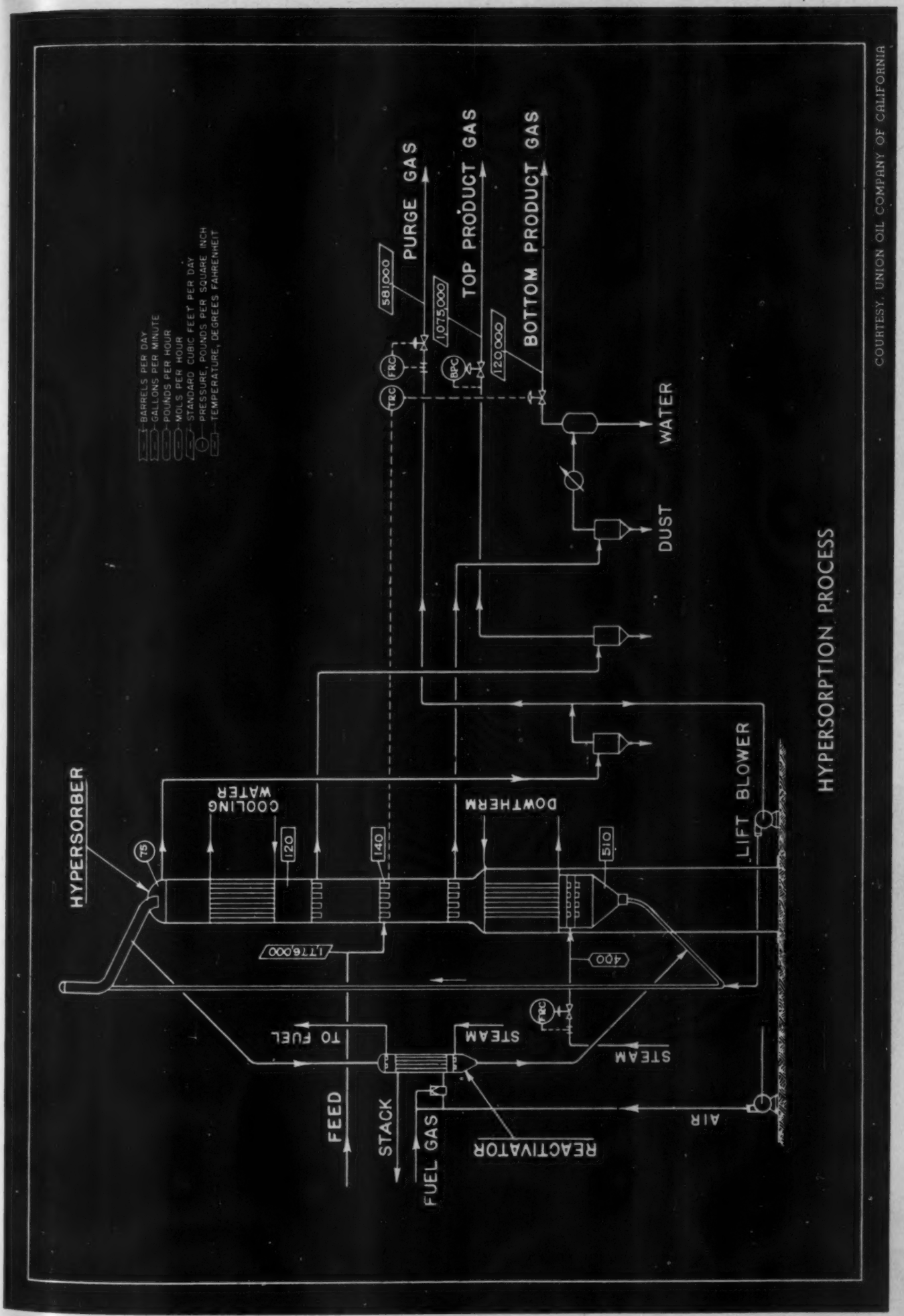
A charcoal regeneration unit is provided to act in parallel with the Hypersorber for the purpose of carrying out high-temperature steam treatment of the charcoal to maintain its activity. If such regeneration were not provided, the activity would slowly drop off due to the accumulation of heavy residual compounds on the charcoal.

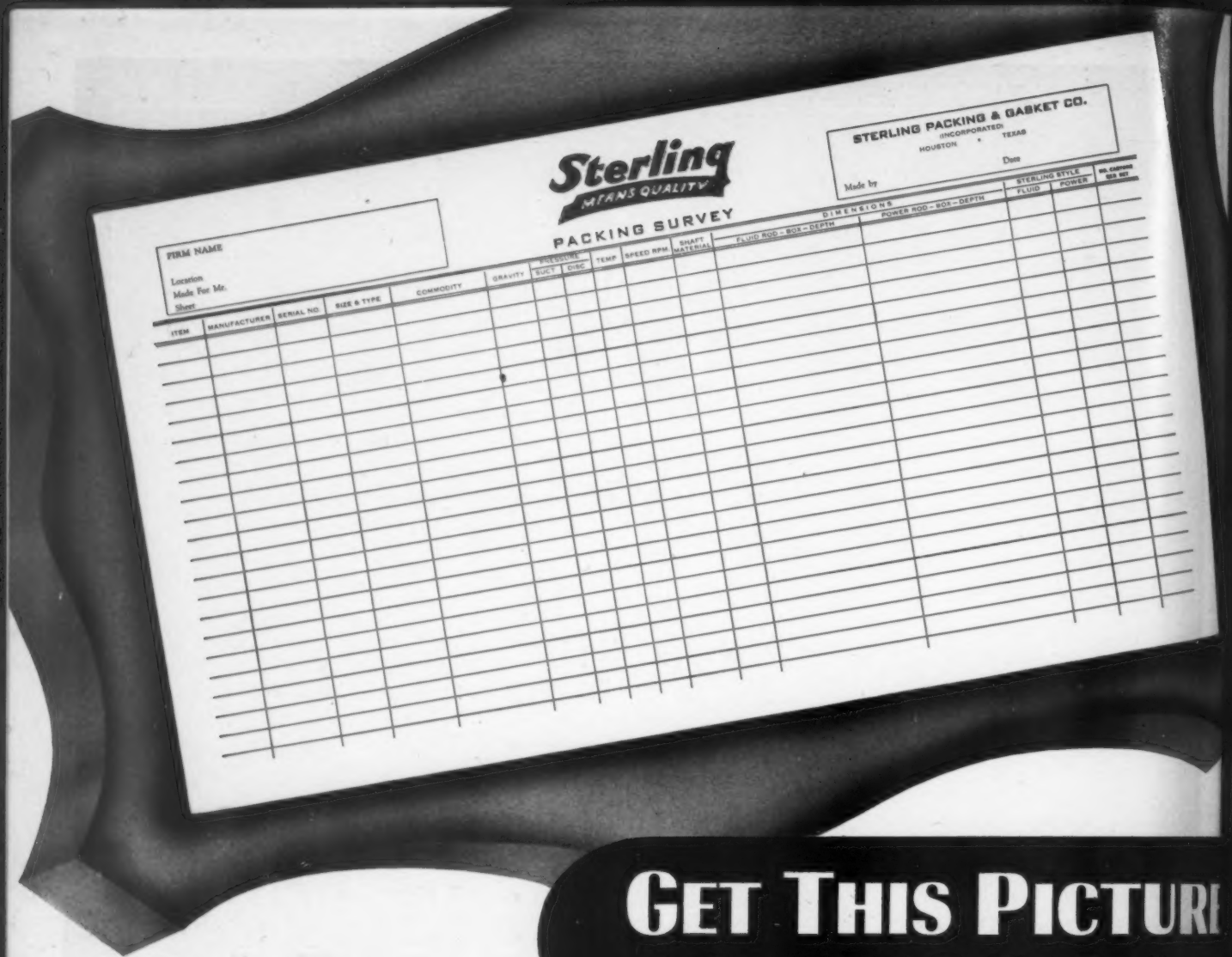
The charcoal is transported to the top of the tower by means of an elevator or gas-lift system and is dropped into the reserve charcoal hopper at the top of the unit. The hot charcoal passes down from the hopper into the cooling section consisting of a tube-and-shell exchanger wherein the charcoal is cooled to the desired temperature for adsorption.

A part of the top product gas stream from the Hypersorber passes up through the cooler and hot charcoal hopper serving to remove any residual water and steam remaining on the charcoal from the stripping operation.

Important to the operation of a Hypersorption unit is the automatic temperature controller employed to maintain a close separation of the components in the unit. A substantial temperature break exists in the adsorber corresponding to the variation in the gas composition in the Hypersorber bed. By controlling this break to a point somewhat below the side-cut disengaging section it is possible to maintain a very close control of the separation process. The temperature controller operates the make-gas control valve which determines the flow of the stripped components to production or reflux. A similar setup controls the side-cut production and purity.

Among the applications for the Hypersorption process are the recovery of propane and heavier hydrocarbons from natural gas or absorption plant dry gas, the separation of hydrogen from methane in hydrogenation and hydroforming plants, and the separation of the ethylene and ethane from hydrogen and methane as well as the heavier constituents of the gas streams in ethylene plants. In this latter application the Hypersorber, among other things, takes the place of the so-called high-pressure, low-temperature demethanizer in the conventional ethylene plant. Hypersorption also is useful for recovery of helium, purification of carbon dioxide, etc. Detailed information on specific problems can be obtained from Union Oil Company of California, Los Angeles.


HYPERSORPTION PROCESS
COURTESY, UNION OIL COMPANY OF CALIFORNIA
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
PURGE GAS
581,000
TOP PRODUCT GAS
1,075,000
BOTTOM PRODUCT GAS
120,000
TRC
FRC
BPC
WATER
DUST
LIFT BLOWER
DOWTHERM
COOLING WATER
HYPERSORBER
75
120
140
510
1,176,000
400
TO FUEL
STEAM
STEAM
AIR
FEED
STACK
FUEL GAS
REACTIVATOR



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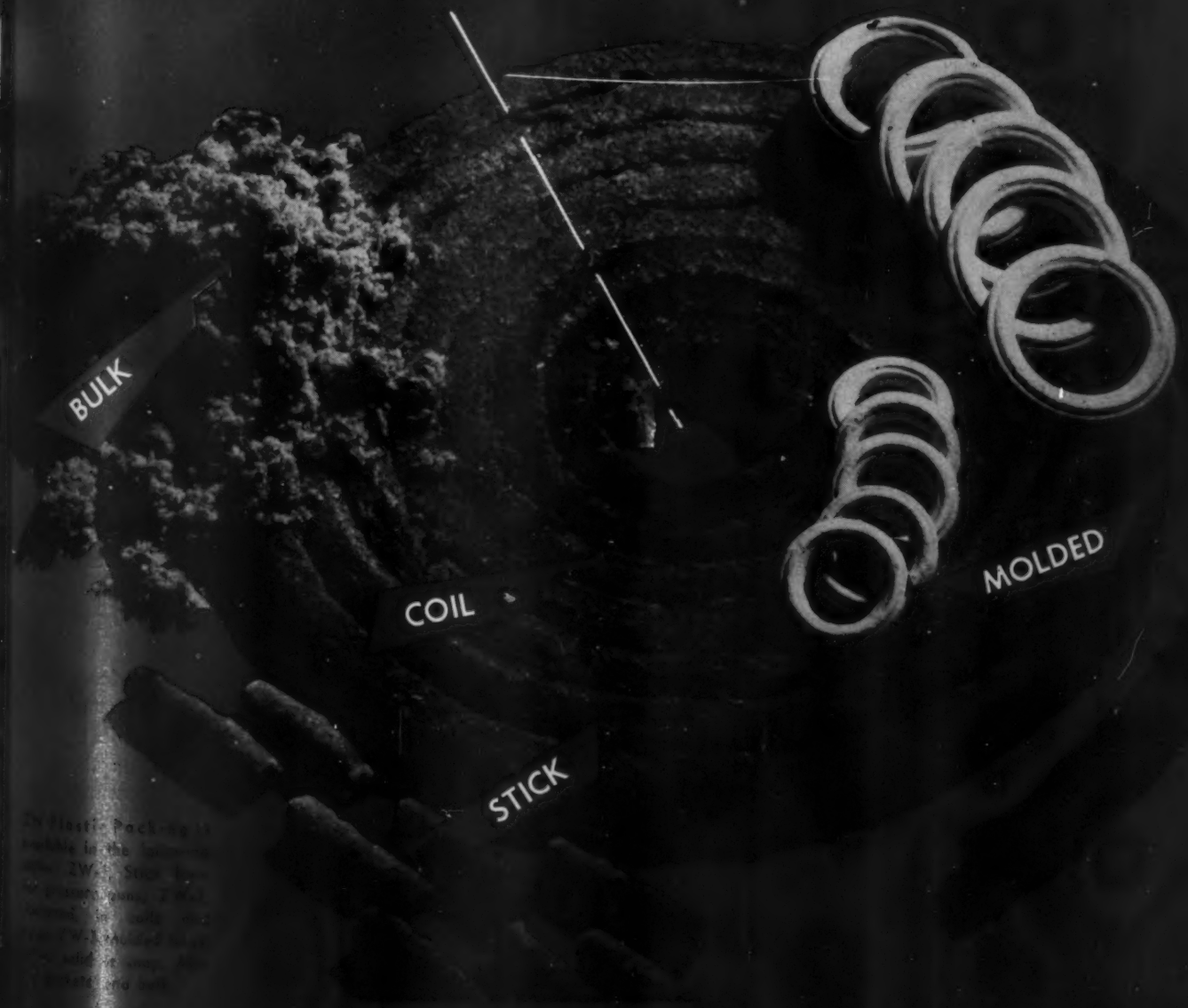

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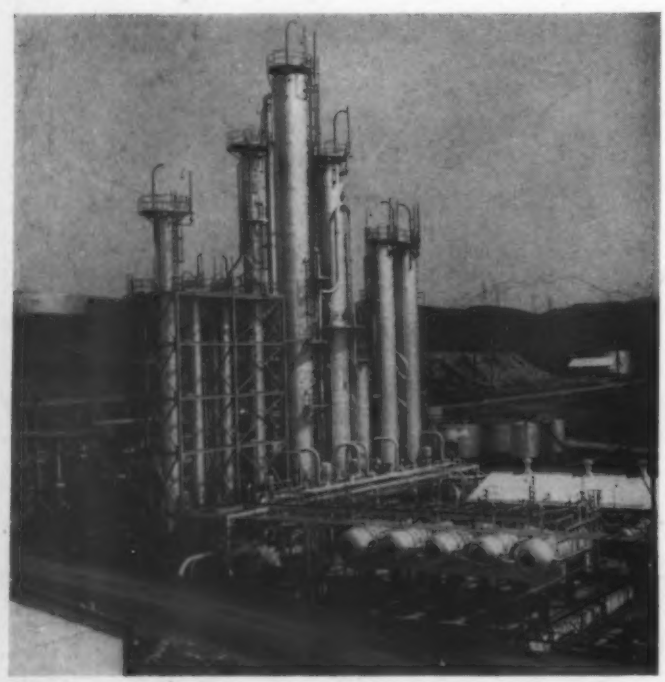
Azeotropic Fractionation Process

UNION OIL COMPANY OF CALIFORNIA

AZEOTROPIC fractionation is a special type of distillation process adapted to the separation of a mixture of compounds that can not be readily segregated by the more conventional distillation procedures. Such mixtures may be composed of either (1) components having essentially the same boiling points such as heptane, boiling point 98.5° C., and methyl cyclohexane, boiling point 100.9° C., or (2) components that form azeotropes among themselves such as cyclohexane and benzene which give a constant boiling mixture (azeotrope) distilling at 77.7° C. The usual petroleum fractions encountered in normal refinery practice represent a combination of both of the above conditions. In many cases, essentially pure compounds can be recovered from such mixtures by adding a carefully chosen material, termed an azeotrope former, which will alter the relative vapor pressures of the components in such a manner that upon distillation one of the components is obtained separately from the others.

The production of nitration-grade toluene from petroleum fractions, an operation which was conducted on a large scale during the war, is a typical example of the advantages of azeotropic fractionation. Toluene occurs naturally in the distillates from certain crudes and the quantity can be considerably increased by the catalytic treatment of such distillates. However, even under the best of conditions when applying conventional distillation procedures, the maximum toluene concentration which can be obtained is the order of only 70-90 percent, depending upon the recovery desired. On the other hand, by adding a mixture of methyl ethyl ketone and water to these toluene-containing distillates, it is a relatively simple matter to recover toluene of 99+ percent purity with yields in excess of 95 percent. The accompanying flow diagram, detailing the recovery of toluene from petroleum fractions, is illustrative of azeotropic-fractionation technique generally, and was based on a commercial plant having an on-stream efficiency of 98 percent.



Referring to the flow diagram, the toluene-containing fraction and the azeotrope former, a mixture of methyl ethyl ketone (MEK) and water, are introduced into the azeotroping column. The constant-boiling mixtures of non-aromatic hydrocarbons, water and MEK pass overhead, giving a bottoms of substantially pure toluene. However, in order to meet Army specifications for nitration-grade toluene, a light acid treatment is required and the acid-treated material then is rerun in a finishing column to give a product of the desired quality. The overall recovery of toluene, based on the toluene present in the feed stock, is above 95 percent.

The overhead product from the azeotroping column is washed with water to recover the MEK. The dilute MEK solution from the extraction tower passes to a recovery column where the MEK-water mixture used as the azeotrope former is distilled overhead and recycled to the azeotroping column. The water from the bottom of the recovery column recycles directly to the extraction column. The initial charge and make-up of this azeotrope former constitute a very small portion of the investment and operating expense of the plant due to the low cost of the MEK and the fact that essentially 100 percent recovery is effected in plant operations.

While the azeotropic process has been described for the segregation of toluene from a petroleum fraction the same general procedure is applicable to the treatment of many other mixtures such as the recovery of benzene and xylenes from petroleum fractions, the separation of aromatic hydrocarbons from thiophenes, paraffins from naphthenes, alcohols from ketones, etc. Detailed information on specific problems can be obtained from Union Oil Company of California, Los Angeles.

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TOLUENE 99% +
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30
NONAROMATICS
364
WATER
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5720
MEK-WATER EXTRACT
AZEOTROPE RECOVERY COLUMN
30
WATER
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ACID TREATMENT
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WATER WASHING
FINISHING COLUMN
20
NITRATION GRADE TOLUENE
85
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HIGH BOILING POLYMERS, ETC.
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
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PRESSURE, POUNDS PER SQUARE INCH
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Petrochemical Processes

IN THE earlier part of the century, chemists interested in petroleum technology predicted that the many hydrocarbons available from petroleum and natural gas would provide the basis for a new organic chemical industry yielding new substances in great variety and of extreme importance to civilization. A serious retarding influence in the development of this new chemical industry existed for many years in the manifold problems of separating from complex mixtures individual chemicals with a high degree of purity, and in commercial quantities. The development of efficient methods of fractionation for the lower boiling hydrocarbons, and fractional solutions in a large number of solvents for isolating the higher boiling compounds, and other technical abilities, have finally initiated the new petrochemicals industry on a quite extensive scale. This relatively new field is presently undergoing growth and expansion so rapid and vast that it is almost without precedent.

The large quantities of hydrocarbons available in petroleum and natural gas constitute a relatively low cost source of raw materials for organic synthesis. The petroleum companies, realizing the value of many products once discarded as waste, or sold in conjunction with products of comparatively low value, are entering the field of organic chemistry. They are beginning to supply old line chemical companies with large volumes of basic organic chemicals for further manufacturing operations. As the manufacture of chemicals from oil and natural gas increases the petroleum companies undoubtedly will engage more heavily in this field.

One aspect of the situation is of great importance. The manufacture of finished organic chemicals varies in considerable degree from the operating techniques employed in the basic oil industry, the liquid fuels industry. Chemicals manufacture deals in comparatively small quantities, and higher investments in both plant and research are required. A higher degree of technical skill and operating control is necessary and plant obsolescence is higher. As a result of these factors, the petroleum companies are finding the manufacture of intermediate chemicals to serve as raw materials for further processing a rather more attractive field, and one well suited to the equipment and operating methods of the refineries and natural gasoline plants. Chemical companies are following this trend in the division of manufacturing operations, too, and are locating plants adjacent or convenient to refineries. The light hydrocarbons which constitute feed stock for many chemical synthesis processes are not easily transported for any considerable distances, and thus greatly influence plant location near the available feed source.

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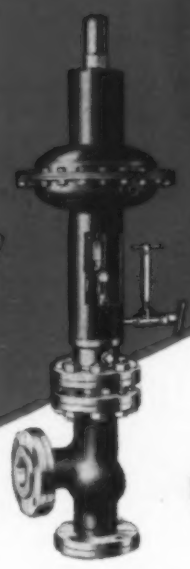

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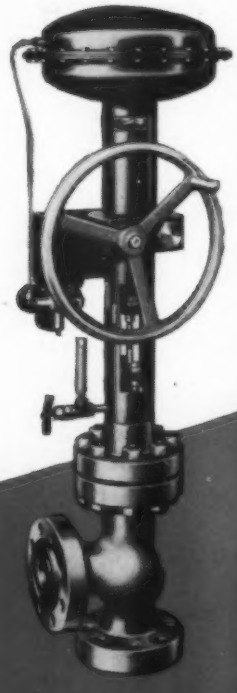

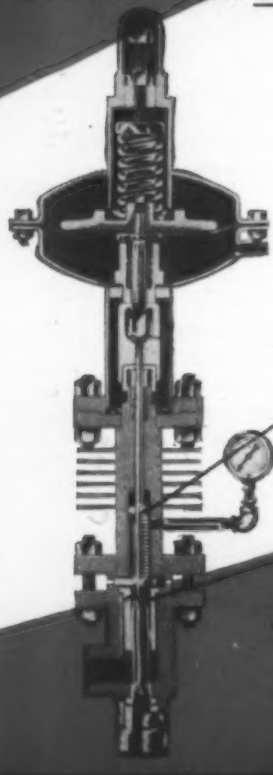

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Anhydrous Hydrogen Chloride Process

THE GIRDLER CORPORATION

THIS process of producing anhydrous hydrogen chloride gas was developed by the Hooker Electrochemical Company and is offered through The Girdler Corporation of Louisville, Kentucky, as licensing agent. Synthetic hydrogen chloride will be found economical when a pure dry gas is needed. It finds many applications in plastics and chemical fields. Chemically pure hydrochloric acid may also be produced by absorbing the gas in distilled water, but it is not generally competitive in price with commercially available muriatic acid.

The basic reaction in the Hooker process is:

$$H_2 + Cl_2 \rightarrow 2\ HCl$$

which is carried out in the water cooled burner vessel indicated on the flow sheet. Essentially, all of the chlorine is reacted and 97 percent or more of the hydrogen. The purity of the finished product depends entirely on the purity of the initial reactants. A typical analysis of the finished gas is as follows:

Hydrogen Chloride	99.00 percent
Chlorine	None
Hydrogen	0.64
Nitrogen	0.20
Water Vapor	0.10
Carbon Dioxide	0.05
Carbon Monoxide	0.01
Total	100.00

Hydrogen produced in a Hygirtol hydrogen plant (described elsewhere in this section) or from electrolytic cells may be used readily. Chlorine may be obtained from tank cars and subsequently vaporized, or from electrolytic cells.

The raw material requirements for producing one ton (2,000 lbs.) of anhydrous hydrogen chloride are approximately as follows:

Chlorine	1,950 pounds
Hydrogen	10,800 SCF*

(* Standard Cubic feet measured at 60° F and 1 Atm.) These should be dry and available at a minimum of 5 psig.

The utility requirements per ton are as follows:

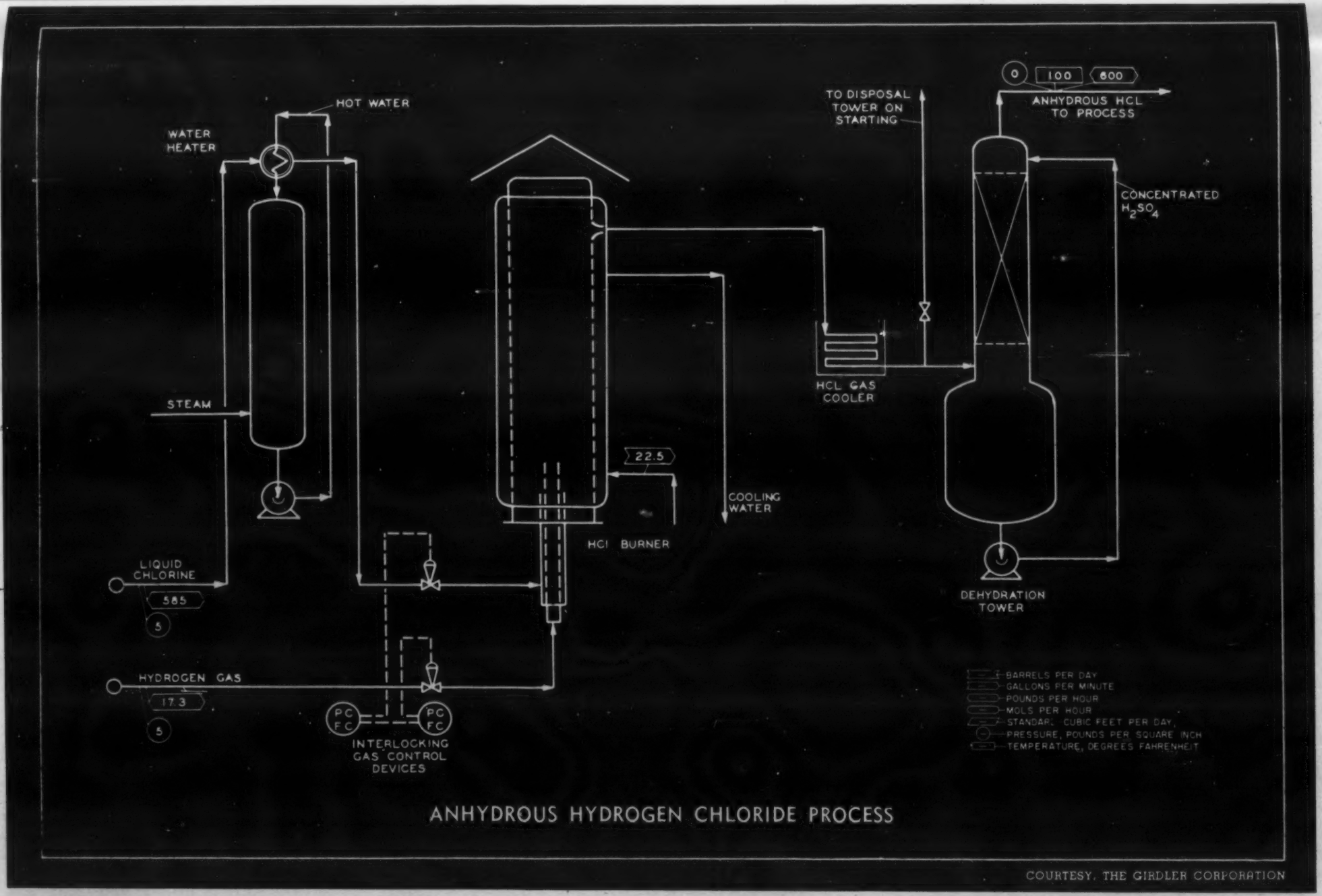
Cooling Water	4000 - 45000 gallons
Power	Nominal
Steam	Space heating only
Labor	1/3 manhour/hour
Misc. reagents for drying, etc.	Nominal

As the hydrogen and chlorine must be metered carefully to avoid the forming of explosive mixtures, specially engineered control devices are employed for this purpose, making the plant essentially automatic in operation. The plant is completely equipped with safety devices to prevent difficulties due to hydrogen, chlorine, or water failures.

Following the burner the gases must be cooled and dried and any remaining traces of chlorine removed. This pure gas is then available at essentially atmospheric pressure and may be used as such or compressed to the desired service pressure. A vent gas absorbing system is used to dispose of HCl and Cl_2 vapors on startups.

The advantages of this method of producing anhydrous hydrogen chloride over other industrial methods are as follows: (a) Low capital investment required per ton of capacity, (b) Low maintenance and operating cost, (c) High purity gas produced directly, (d) Safer operation (e) Small capacity installations possible and economically feasible, (f) Plant output may be readily varied to suit consumption, (g) No by-product disposal problem.

This process has been operated on a full commercial scale for more than five years by Hooker and others.



ANHYDROUS HYDROGEN CHLORIDE PROCESS

COURTESY, THE GIRDLER CORPORATION

Ammonia Synthesis Process

LION OIL COMPANY

CONVERSION of natural gas into ammonia came into wide application during the war period as a source of explosive ingredients. One of these plants, Ozark Ordnance Works was operated by Lion Oil Company, El Dorado, Arkansas, and has since become a peacetime activity of the company, which now produces anhydrous ammonia, aqua ammonia, ammonia-ammonium nitrate solutions as well as ammonium nitrate fertilizer, from natural gas, air and water.

This process description and the accompanying flow diagram were provided by the Chemical Division of the company and represent expanded operations of the war-built plant. From a product viewpoint the process consists of three parts: manufacture of anhydrous ammonia, manufacture of nitric acid, and the finishing of fertilizer.

Steps in the manufacture of anhydrous ammonia include preparation of the raw (suction) gas at the reform plant, purification of the suction gas, and fixation of the purified hydrogen-nitrogen mixture in the synthesis operation to form ammonia. Reforming of natural gas is accomplished by passing a mixture of gas and steam downward through tubes and over a nickel catalyst at 725° C. Air is added to furnish nitrogen, the oxygen being consumed by hydrogen, and simultaneously a portion of the residual methane is converted into hydrogen by secondary reforming. The water-gas reaction carried out in the CO converter in the presence of iron oxide catalyst then accounts for the remainder of the hydrogen produced in the reform step.

Suction gas containing approximately 20 percent impurities (CO_2, O_2 and CO) is first scrubbed with water at 225 pounds pressure to remove most of the carbon dioxide. Then the gas is scrubbed with ammoniacal copper ammonium formate-carbonate solution and then 5 percent caustic at 1800 pounds pressure for removal of remaining carbon dioxide, all of the oxygen, and all but a trace (10 ppm) of carbon monoxide.

In the synthesis loop the purified process gas (make-up) and recycle gas both enter the oil filter at 5000 pounds pressure, pass to the cold exchanger, thence to the secondary or ammonia-cooled condenser, back to the cold exchanger, and then to the ammonia converter where, by exchange with the hot exit gases, the inlet gas is heated and enters the catalyst bed at 450° C. Passing through the catalyst, fixation of the hydrogen and nitrogen takes place at 5000 pounds pressure and a hot-spot temperature of 525° C. Only a portion of the hydrogen-nitrogen is converted, the converter inlet gas containing 2.5 percent ammonia and the outlet 12.5 percent. An iron catalyst promoted with aluminum and potassium oxides is used in the conversion of the hydrogen-nitrogen mixture to ammonia. The gas passes from the converter to the primary or water-cooled condenser, from which it enters the primary separator, and then goes to the circulator for recompression, and delivery of the recycle gas to the oil filter. Anhydrous liquid ammonia is continuously drawn off the cold exchanger and primary separator.

The process for manufacturing nitric acid comprises three steps: ammonia oxidation, secondary oxidation and absorption. In the first step an ammonia-air mixture at 60 pounds pressure is passed over a hot platinum-rhodium gauze, maintained at 900° C., and burned to nitric oxide (NO). The gases are cooled and then pass to the oxidation tower where the introduction of secondary air produces the nitrogen dioxide (NO_2). This material then is absorbed in water by means of a system of absorption towers, circulating pumps, and trickle coolers to form 56 to 57 percent nitric acid.

In the manufacture of the ammonium nitrate fertilizer nitric acid (56-57 percent) first is neutralized with anhydrous ammonia to form an 83-percent ammonium nitrate solution. This solution is concentrated in a vacuum evaporator and the melt charged to a pelletizing tower. The pelleted material is removed, dried, cooled, and coated to produce the desired storage and drillability qualities. The latest design which industry affords is incorporated into the process and a superior product is obtained.

A unique feature in the design of the Lion plant is that natural gas is used throughout as the prime source of power. Such includes its use as fuel in the primary reform furnaces, as fuel at the steam plant, and as fuel for driving gas engines. These gas engines, 54 in number, are housed in the compressor building and represent nearly 60,000 horsepower. All the engines are 1000-horsepower each with the exception of 11 1000 K.W. engines used on electric-power generation. Twenty-four of the engines are used on process gas compression, 5 on refrigeration, 6 on air compression for nitric acid, and 8 for driving water pumps in the water scrubbing operation.

The chemical plant is operated as a division of Lion Oil Company.

AMMONIA-NITRIC ACID PROCESS
COURTESY, LION OIL COMPANY, CHEMICAL DIVISION
GAS REFORMING PLANT
AMMONIA PLANT

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R-S hand-controlled valves are fully closed from full open position by the movement of a lever through 77½° of arc. Either lever or wheel may also be actuated by a chain.

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Uniform Control in all Positions

R-S Valves are unique in giving uniform control of flow over most of the arc through which the disc travels from full open to full closed position.

More Compact, Lighter, Lower in Cost

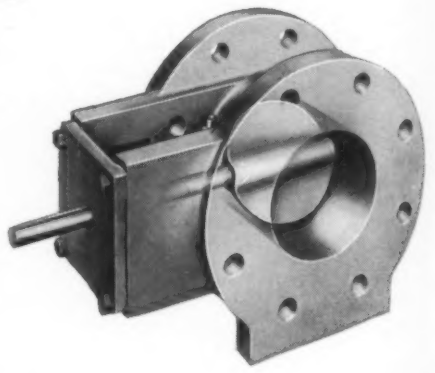
R-S Valve design gives a narrower face-to-face than most types of design. The valve is lighter and requires less support. Less metal is required for its construction; therefore less machining, resulting in lower cost.

Simpler . . . Self-cleaning . . . Non-turbulence Reduces Wear

R-S Valve design offers no pockets to capture sediment, no change of flow-direction to create turbulence. Freedom from turbulence reduces erosion and cavitation and therefore adds to the life of the valve.

A Principle

Every R-S Valve is designed and built for extremely rugged service, and provided with such safety-factors that it will give complete satisfaction in the service for which it is designed.



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Styrene Process

THE DOW CHEMICAL COMPANY

THE Dow Styrene Process was developed during the early 1930's at The Dow Chemical Company's Research Laboratories in Midland, Michigan, as a step in the manufacture of the plastic polystyrene. The wartime synthetic rubber program resulted in construction of four plants which supplied styrene for GR-S rubber. Although the process was licensed during the emergency, it is not generally available.

The process in brief is the catalytic dehydrogenation of ethylbenzene to styrene at a conversion per pass of about 37 percent and an ultimate yield of 90 percent, with subsequent vacuum distillation of the crude styrene.

Since reduced pressure favors the dehydrogenation reaction, steam is fed at a rate of 2.6 pounds per pound of ethylbenzene. About 90 percent of this steam quantity is heat-exchanged to a temperature of 725° F. before entering the superheating furnace, which produces steam at 1310° F. The remaining 10 percent of the steam is mixed with ethylbenzene before being passed through a vaporizer, attaining a temperature of 320° F. before heat exchange with hot reactor product. At 968° F. the ethylbenzene vapors meet superheated steam at 1310° F. The resulting temperature at the bottom of the catalyst bed is 1166° F. This temperature point is used for controlling the reaction system. Reactor product at 1049° F. is cooled by heat exchange, first with incoming ethylbenzene, and then with steam. From this point a spray-type desuperheater lowers the crude product temperature to 221° F. and condenses tars formed at high temperatures. Vapors then enter the final condenser, where steam, styrene, ethylbenzene, benzene, toluene, and small amounts of tar are liquefied. Vent gasses containing hydrogen, carbon monoxide, carbon dioxide, methane, ethane, plus some of the aromatics pass overhead to a refrigerated recovery system.

The materials condensed go to a gravity separator, where the hydrocarbons are decanted from the water phase. After further settling, part of the water is recycled to the sprays of the desuperheater, and the rest is discharged to a disposal system. The hydrocarbons pass to another settler, where insoluble tar can drop out along with entrained water. Crude styrene then is pumped to storage. An average composition for this stream by weight is:

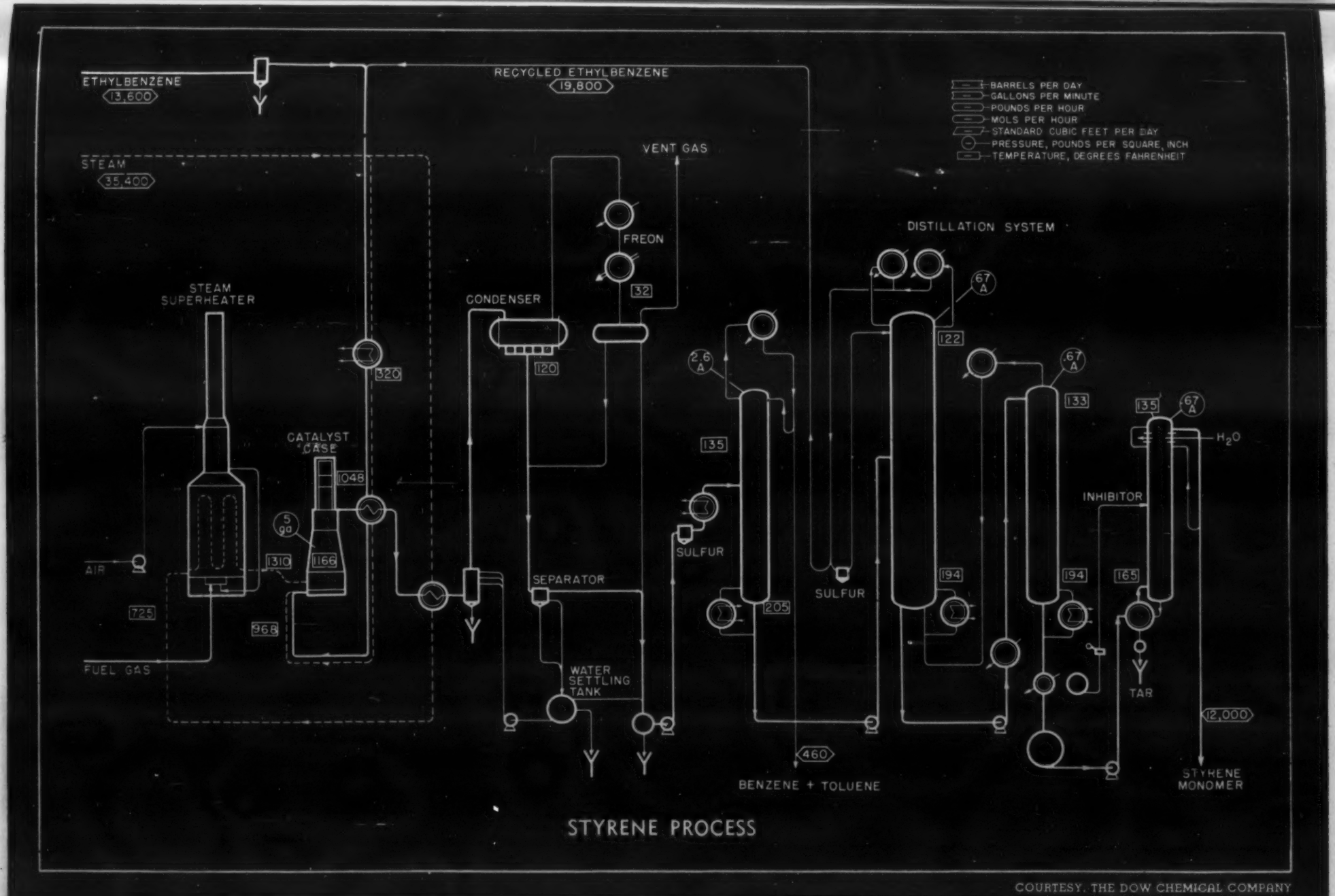
Styrene	37.0 percent
Ethylbenzene	61.1
Toluene	1.1
Benzene	0.6
Tar	0.2

The final step is the purification of the crude dehydrogenated material containing about 37 percent styrene. The fractionation requirements are rather strict since the normal boiling points of ethylbenzene and styrene differ by only 48° F. Other factors complicate the distillation step to an even greater extent. Distillation of styrene at its atmospheric boiling point (293°F.) is impossible because of rapid formation of polystyrene. At 30 mm. Hg. absolute, the boiling point of styrene is 129° F., and the polymerization rate is less than 0.1 percent per hour. Running under a vacuum, then, is necessary. Even though a low distillation temperature is achieved, the normal irregularities of operation make further precautions necessary. Very small quantities of certain foreign materials catalyze the polymerization reaction greatly. To guard against such possibilities, elemental sulfur is dissolved in styrene to act as a polymerization inhibitor. Only by a combination of vacuum operation, suitable inhibition of styrene, and special column design can styrene be distilled successfully. Relatively trouble-free operation can be had by distilling inhibited styrene in a system which never allows the concentrated monomer to exceed 194° F.

In the following the flow further, crude styrene passes through a pot containing sulfur. Enough sulfur is dissolved to act as a polymerization inhibitor. After pre-heating, the feed enters the 30-plate benzene-toluene column where benzene and toluene pass overhead at 135° F. The top of this column operates at 175 mm. Hg., and reflux ratio of 12 to 1 is employed. Since styrene is still diluted with ethylbenzene and tar, a temperature of 205° F. is allowable at the bottom. Sulfur passes down the column and effectively inhibits the stripping zone. Benzene and toluene from all four finishing blocks then pass to a benzene column of standard construction operating at atmospheric pressure. It contains 40 plates and operates with a reflux ratio of 3 to 1. Benzene, thus purified to less than 0.1 percent toluene content, is recycled to the ethylbenzene plant for alkylation. The bottoms from this column contain toluene and traces of ethylbenzene. Toluene with a 34° F. boiling range is obtained as product from the 35-plate toluene column. Ethylbenzene, styrene, and tar then pass to the primary ethylbenzene column. At this stage comes the rather difficult separation of ethylbenzene from styrene. A single column of 70 plates would give a bottom pressure of 225 mm. Hg. and a condition where highly concentrated styrene would be boiled at 219° F. This temperature would not be advisable even in the presence of sulfur. To overcome this difficulty the separation is split into two steps: The primary ethylbenzene column, which acts as the top 38 plates, and the secondary ethylbenzene column acting as the bottom 32 plates. Vacuum jets and cooling water keep the top of each at 35 mm. Hg., and thus, the bottom temperatures can be held at a safe level of 194° F. The reboilers of these stills are short-tube vertical units designed for a very low liquid head in tubes. Steam at 5 pounds gauge is on the shell side. Low T's in the reboilers are necessary to prevent high skin temperatures.

The columns are designed for a low pressure drop and many precautions are taken to prevent the inward leakage of air. The upper part of the primary column is protected against polymer formation by the addition of sulfur to ethylbenzene reflux. This primary column takes recycle ethylbenzene containing about 1 percent styrene overhead to the feed of the dehydrogenation step. Bottoms from the primary are pumped through a cooler to the top of the secondary column. The overhead from the secondary column is condensed, then flows by gravity to the reboiler of the primary. Condensation of this stream has proved desirable from the control standpoint. Bottoms from the secondary column are cooled to prevent polymerization and then pass to a batch-still charge tank which is held at 35 mm. Hg. pressure. Packed batch styrene finishing stills are used to remove tar and sulfur from the styrene. Batch operation permits somewhat easier handling of the residue and also gives greater flexibility to the system. A styrene finishing still runs at varying reflux ratio depending on the purity of the overhead product. Product purity is recorded continuously on a strip chart by a specially designed infrared analyzer. Top temperature is 135° F. at a pressure of 35 mm. Hg. Since the upper part of the column operates with little sulfur present, a solution of para-tertiary-butyl catechol in styrene is pumped into the reflux to this still by means of a proportioning pump. Five parts per million of inhibitor are effective in preventing excessive polymer formation. The residue from these columns is drained while hot into dump buckets. This heavy material, consisting of tars formed in the dehydrogenation step, styrene-sulfur compounds, elemental sulfur, polystyrene, and styrene are burned. Before shipment, para-tertiary-butyl catechol is added to bring the concentration to a minimum of 10 ppm. Shipment of the finished material is made in insulated tank cars. Styrene is pumped through a refrigerated cooler to these cars to provide a loading temperature below 68° F. With these precautions, the probability of polymer formation in normal tank-car shipment is exceedingly small.

The process described is highly instrumented. The on-stream time of the units is quite high—98 percent or more.


ETHYLBENZENE
13,600
RECYCLED ETHYLBENZENE
19,800
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE, INCH
TEMPERATURE, DEGREES FAHRENHEIT
STEAM
35,400
VENT GAS
FREON
32
DISTILLATION SYSTEM
STEAM SUPERHEATER
CONDENSER
120
320
67 A
122
2.6 A
133
135
CATALYST CASE
1048
5 gal
310
166
AIR
725
968
FUEL GAS
SEPARATOR
SULFUR
205
194
165
INHIBITOR
H_2O
TAR
WATER SETTLING TANK
460
BENZENE + TOLUENE
12,000
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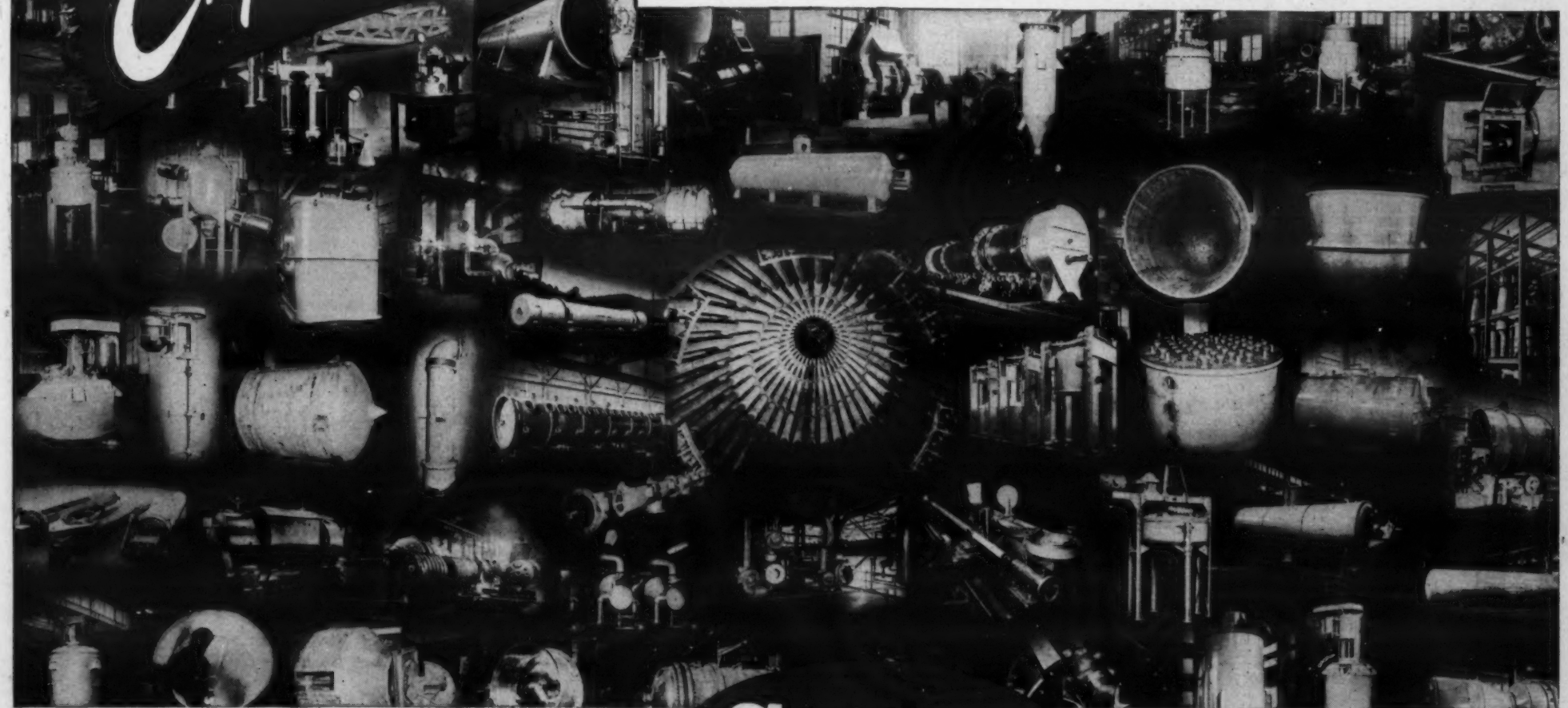

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Ethylene Process

STONE & WEBSTER ENGINEERING CORPORATION

THE two principal sources of ethylene for chemical synthesis are (1) olefin-rich refinery gases and (2) thermal cracking of ethane or propane primarily for the production of ethylene. Most chemical synthesis processes require that the ethylene supplied be relatively pure. In most cases an ethylene concentration in the range 95-99 mol percent is required.

The low temperature fractionation process shown on the accompanying flow sheet is one of the most widely used of the various ethylene separation processes, roughly 75 percent of the ethylene produced in the United States being produced by this process.

The fresh feed, together with recycled ethane and C_3's, is charged to the pyrolysis furnace and heated to 1425 - 1450° F. where optimum conversion to ethylene takes place. The furnace design is such as to give a contact time in the reaction zone equivalent to 0.7 - 1.3 seconds at the coil outlet temperature. The cracked gas mixture is water quenched at the coil outlet to below cracking temperature, and then cooled to about 100° F. in a water cooled quench tower.

The cracked gas is compressed in three or four stages to 550 - 600 pounds per square inch gage. The gas from the last stage of compression is subcooled to about 65° F. to remove the maximum amount of water before introducing it into the dehydrators. Drying of the gas to a water dew point of approximately minus 100° F. is effected by passing the gas through a bed of activated bauxite or alumina.

The dried gas is cooled and partially liquified by heat exchange and refrigeration, then introduced into the demethanizer tower. Hydrogen and methane are taken overhead in the demethanizer together with about 3 to 5 mol percent of C_2's required to saturate the overhead stream at 550 pounds per square inch absolute and a reflux condenser temperature of minus 130 to minus 140°F. Because of the high relative volatility of the key components methane and ethylene, only a very low reflux ratio is required in the demethanizer. Reflux is produced in a run-back type condenser refrigerated by the evaporation of ethylene.

Subsequent processing of the ethylene and heavier components in the demethanizer bottoms varies with the composition of the bottoms and number and purity of products required. In the process shown on the flow sheet the bottoms are fed to an ethylene tower where high purity ethylene is removed as an overhead product. Reflux is produced at 0 to minus 15° F. in a condenser refrigerated with propane or ammonia. Ethylene tower bottoms are fed to a recycle tower where ethane and C_3's are taken overhead and recycled to the cracking furnace. The recycle tower bottoms consisting of C_4's and heavier usually are not worth further processing and are discarded to fuel. When cracking distillate stock, however, this fraction is a worthwhile source of aromatic distillate.

For simplicity, the refrigeration system has not been shown. A cascade type system is employed, the ethylene being condensed by the ammonia or propane refrigerant at about 0° F., this in turn being condensed with cooling water. In the large capacity ethylene plants built in the last few years, it has been feasible to incorporate into the refrigeration system refinements that have increased the efficiency of the refrigeration cycle with resulting savings in power consumption.

Operation of the plant is continuous, the solid adsorbent type dehydrators reducing the water content of the gas to a low enough value that no ice formation is experienced in the low temperature equipment. Low operating temperatures throughout the recovery unit keep polymer formation and exchanger fouling to a minimum.

ETHYLENE PROCESS
COURTESY, STONE & WEBSTER ENGINEERING CORPORATION
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
ETHANE-PROPANE FEED
RECYCLE
PYROLYSIS FURNACE
425
QUENCH TOWER
20
5.3
QUENCH H2O ACCUMULATOR
DEHYDRATORS
520
10
PYROLYSIS GAS COMPRESSOR
METHANE TOWER
-135
500
ETHYLENE TOWER
8
400
RECYCLE TOWER
55
150
RECYCLE ACCUMULATOR
ETHYLENE PRODUCT
RESIDUE GAS
BUTANES AND HEAVIER

Ethylene Process

THE LUMMUS COMPANY

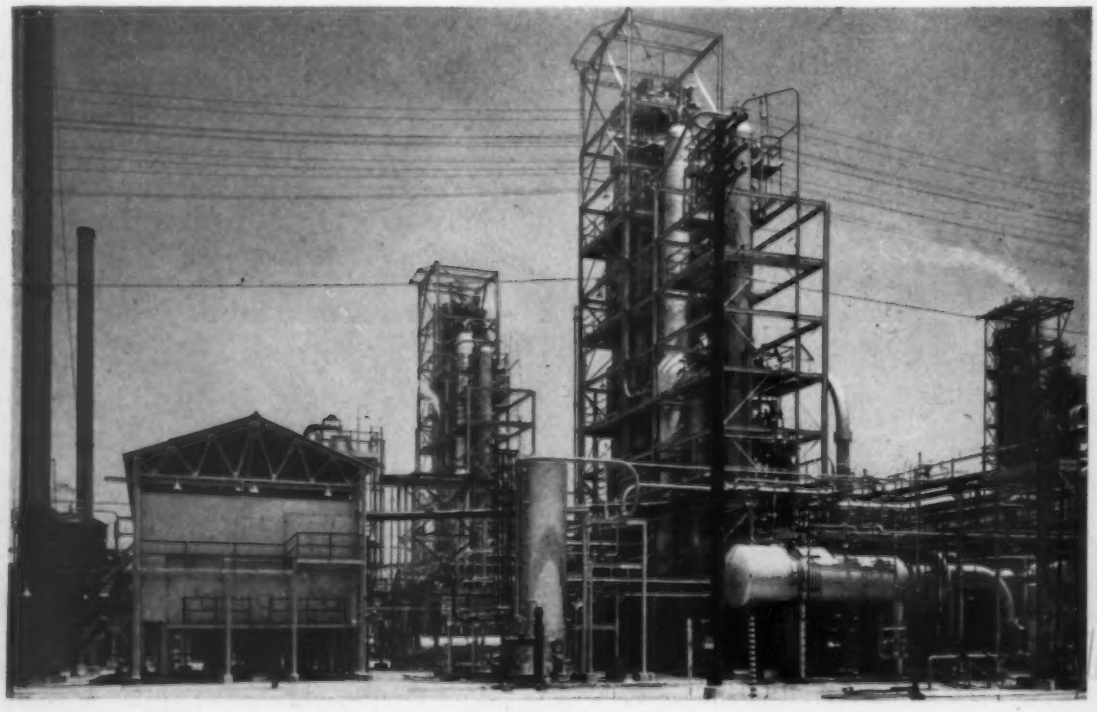
THE great expansion of the petroleum-chemical industry during the past few years has been reflected in the growth of ethylene manufacture and recovery from about 150,000 pounds per day in 1940 to about 600,000 pounds per day in 1947, with more than one million pounds per day scheduled by 1948. While other petroleum source materials, such a benzene, toluene, and the light olefins heavier than ethylene contribute significantly to the tonnage of petroleum chemicals today, ethylene has become probably the most important of these source materials.

Ethylene may be produced by the pyrolysis of ethane, propane or heavier materials, including reduced crudes, by short-time high-temperature pyrolysis and of course may be recovered from oil-refinery gases where it is the by-product of relatively low-temperature thermal and catalytic-cracking operations. The high-temperature cracking may be conducted in the presence of hot granular refractory or may be accomplished in tubular furnaces. The most important source material for ethylene pyrolysis at the present time is ethane and propane which may be efficiently processed in tubular heaters at temperatures in the range of 1300-1400° F. The flow sheet opposite represents a typical plant designed to produce ethylene by the pyrolysis of ethane, propylene and propane, and to recover ethylene directly from the raw charge gas.

The fractionating absorber is the most important unit in the plant in that it makes the separation between methane and ethylene. It is necessary to avoid loss of ethylene, but on the other hand, retention of too much methane reduces the purity of the ethylene below the desired value since generally no further demethanization is employed. A portion of the aromatic distillate produced in the process is circulated over the absorber at a rate of approximately 3.0-4.0 pounds per pound of feed, at which rate only moderate heat removal in the intercoolers is required in order to recover approximately 99 percent of the ethylene entering in the feed and reject all of the hydrogen and approximately 96 percent of the methane. The rich oil from the absorber flows to the deethanizer where the split is made between ethane and propylene. Inasmuch as both the ethane and the propane-propylene are recycled through the pyrolysis coils, the separation between propylene and ethane need not be sharp, and a low reflux ratio may be used. In this tower it is only necessary to avoid loss of ethylene in the bottoms. The overhead from this tower flows to the ethylene fractionator where the split is made between ethylene and ethane. Recovery of practically 100 percent of the ethylene in the feed to this tower with purities up to 95 percent can be made readily.

Aromatic materials removed from the bottom of the depropanizer are used for absorber lean oil as required, with the balance being delivered to aromatic-distillate production. A small rerun tower is provided for eliminating heavy polymer which builds up in the aromatic distillate.

In this system refrigeration is required in the absorber, the deethanizer and the ethylene fractionator. However, the refrigeration level is from 0° F. upwards and the refrigeration load is therefore moderate amounting to approximately 18 brake-horsepower for 1000 pounds of net ethylene produced per day, exclusive of the charging stock prefractionation requirements. Driers are provided for reducing the water vapor dew point of the absorber charge to less than —40° F., in order to eliminate the formation of ice and hydrates over prolonged periods of operation.



LUMMUS ETHYLENE UNIT
LEGEND
WATER COOLED HEAT EXCHANGER
REFRIGERATED HEAT EXCHANGER
ETHYLENE
DEPROPANIZER
RERUN TOWER
ETHYLENE FRACTIONATOR
DEETHANIZER
HEATERS
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ABSORBER
DRIERS
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COURTESY, THE LUMMUS COMPANY

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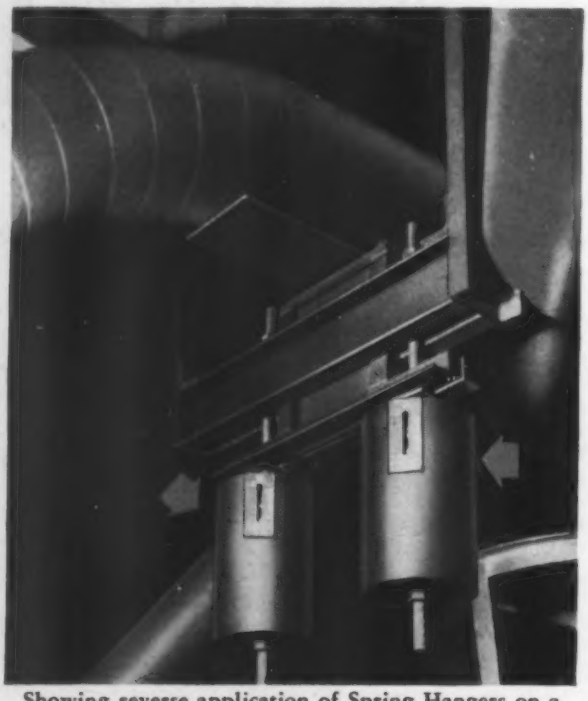
Selecting hangers for piping subject to thermal movement is completely simplified. Grinnell Pre-Engineered Spring Hangers are available in stock sizes for any load requirement.

Simply compute the load and order the proper size Grinnell Pre-Engineered Spring Hanger as indicated in the Capacity Table appearing in descriptive catalog. 14 sizes with load range from 84 lbs. to 4,700 lbs.

Flexible suspension of piping in one of world's largest catalytic cracking plants.



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LOOK AT THESE FEATURES . . .

- **The maximum variation in supporting force per ½" of deflection is 10½% of rated capacity—in all sizes.**
- **Guides prevent contact of coils with casing wall or hanger rod and assure continuous alignment and concentric loading of spring.**
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- **Precompression* assures operation of spring within its proper working range where variation in supporting force is at a minimum.**
- **All-steel welded construction meets pressure piping code.**
- **14 sizes available from stock—load range from 84 lbs. to 4700 lbs.**
- **Easy selection of proper sizes from simple capacity table.**
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- **Unique swivel coupling provides adjustment and eliminates turnbuckle.**

***Precompression is a patented feature**

Write for descriptive folder on Pre-Engineered Spring Hangers Fig. 268.

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4. Easy Inspection and Maintenance
5. Continuous Operation. No Shut Downs for Servicing.
6. Minimum Investment in Replacement Parts.
7. The Brown Resistance Welded Integrally Bonded Fintube.
8. Easily Removed Rear End Cover Plate.

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● Unlike "bundle" type exchangers, which are custom built for a definite and fixed duty, Brown Fintube Sectional Hairpin Heat Exchangers give operating men almost unlimited flexibility to meet changing requirements. The capacity of the bank can be increased by adding additional sections to the original installation. Capacity can be decreased by valving off one or more of the parallel streams,—or reducing the number of sections in each stream. The entire bank can be changed from one duty, to another and totally different duty, by remanifolding the sections in different series-parallel arrangement. And the bank can even be disassembled, and the sections then reassembled as two or more smaller exchangers on different duties, in different plants.

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Sectional Hairpin

HEAT EXCHANGERS

RESISTANCE WELDED INTEGRALLY BONDED FINTUBES AND FINTUBE HEAT TRANSFER PRODUCTS

Hydrogen Process

THE GIRDLER CORPORATION

A HYGIRTOL* Hydrogen plant for manufacturing high-purity hydrogen from hydrocarbons and steam combines several well established steps into a unified, continuous, low-pressure process for producing hydrogen of purity equal to or better than that produced by any other commercial process. For many uses the hydrogen from a Hygirtol Plant is preferred even to that produced by electrolysis because its oxygen content is so low.

Suitable hydrocarbons such as natural gas, refinery gases, propane, butane and others, are purified to remove sulphur compounds, and then are reacted with steam at elevated temperatures (about 1500° F.) over a nickel catalyst to convert the hydrocarbons to hydrogen, carbon monoxide, and carbon dioxide. By employing suitable catalysts in a furnace of special design, the decomposition of the hydrocarbons is substantially complete so that the hydrogen produced will contain less than 0.1 percent of residual hydrocarbon.

The following chemical reactions occur in the furnace:

$$(1) \quad C_nH_{2n+2} + nH_2O = nCO + (2n+1)\,H_2$$
$$(2) \quad C_nH_{2n+2} + 2nH_2O = nCO_2 + (3n+1)\,H_2$$

Similar reactions occur with unsaturated hydrocarbons such as ethylene and propylene. These reactions are all endothermic, so that heat must be supplied in the furnace to maintain the required temperatures.

The mixture of hydrogen, carbon monoxide, carbon dioxide, and steam from the furnace is cooled to about 700° F. by the further addition of steam and is passed into a converter containing an iron-oxide catalyst where the carbon monoxide reacts with steam to produce carbon dioxide and hydrogen as shown in the following equation:

$$(3) \quad CO + H_2O = CO_2 + H_2$$

This reaction is slightly exothermic, so no heat need be added to the converter. It is a reversible reaction, with about 90 to 95 percent of the carbon monoxide being converted to carbon dioxide under the conditions normally encountered.

The gas mixture from the converter is cooled and the carbon dioxide is removed by the Girbotol Process. The hydrogen is then purified further as required to remove the residual carbon monoxide (about 1 percent) by further carbon monoxide conversion steps followed by the removal of carbon dioxide by the Girbotol Process, or by methanation of the residual carbon monoxide according to the following reaction:

$$(4) \quad CO + 3H_2 = CH_4 + H_2O$$

This reaction is exothermic and is carried out at slightly elevated temperatures in the presence of a catalyst.

The flow diagram on the opposite page illustrates a Hygirtol hydrogen plant producing high-purity hydrogen using propane as process material and fuel. A typical analysis of the hydrogen produced in a plant of this type is as follows:

Analysis of Hygirtol Plant Hydrogen

Component	Percent
Carbon dioxide	0.01
Carbon monoxide	0.00
Methane	0.10
Nitrogen	0.01
Oxygen	0.00
Hydrogen	99.88
	Total 100.00

The material and utility requirements for producing 1000 cubic feet of high-purity hydrogen in this type of plant are given below:

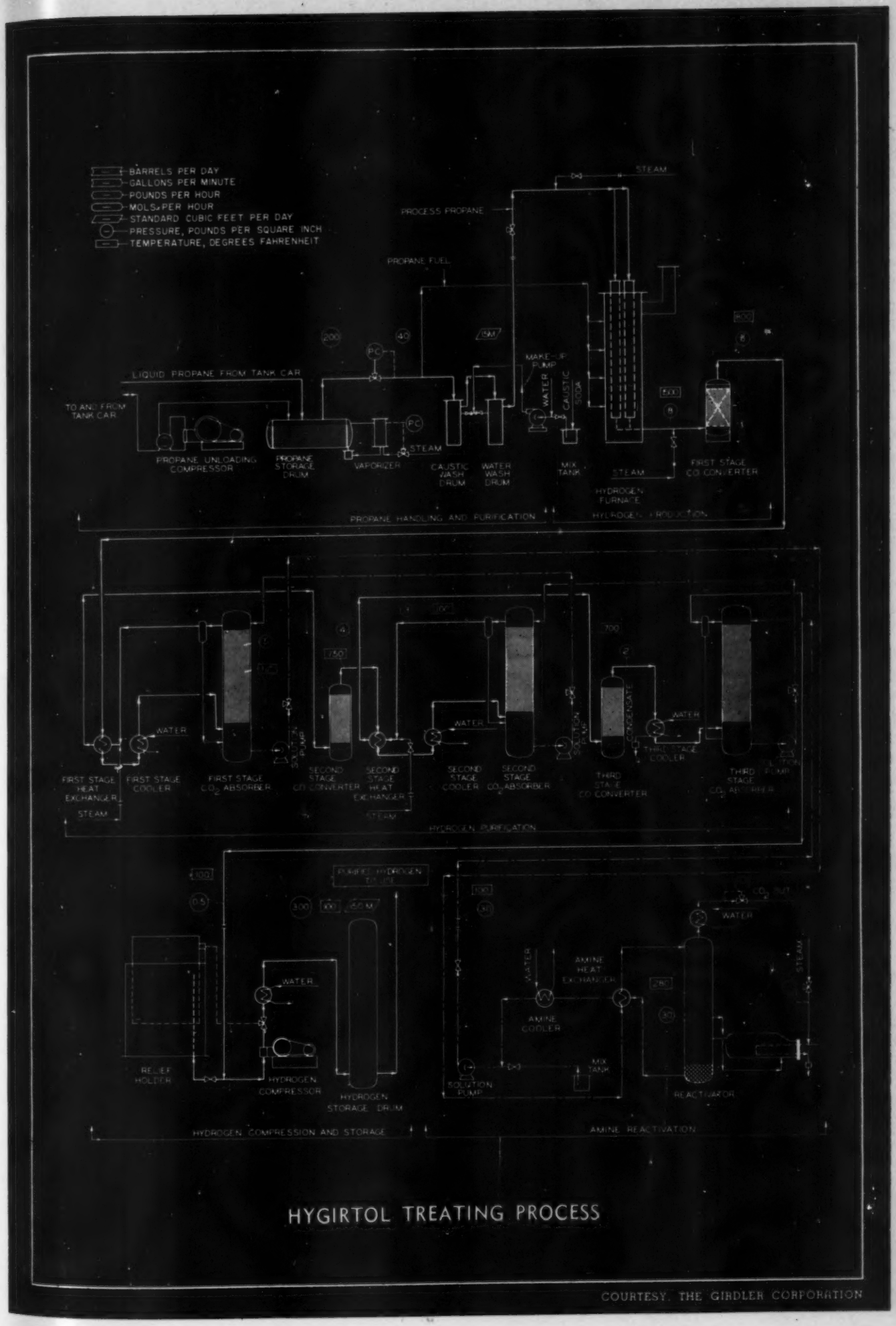
Material and Utility Requirements per 1000 SCF of Hydrogen

Material or Utility	Quantity
Process Material:	
Propane, gallons	2.75
Or Natural gas, cubic feet	250
Fuel (gas or oil) BTU	250,000
Steam pounds	380
Cooling water, gallons (30° F. rise)	1600
Power, KWH	2
Chemicals, cost in cents	2

The Girdler process is continuous, and it can be made completely automatic in operation. The catalysts used are rugged and long-lived, so that replacement costs amount to only one or two cents per thousand cubic feet of hydrogen produced. Only one operator is required per shift, and maintenance is very small.

This information is presented through the courtesy of The Girdler Corporation, Louisville, Kentucky, engineers and builders of Hygirtol hydrogen manufacturing plants.

* "Hygirtol" is a trade mark of The Girdler Corporation.



HYGIRTOL TREATING PROCESS

COURTESY, THE GIRDLER CORPORATION

Hydrocol Process

TEXACO DEVELOPMENT CORPORATION

THE Hydrocol Process, developed by Hydrocarbon Research, Inc., for the production of premium quality gasoline, diesel fuel oil and other products by the reaction of carbon monoxide and hydrogen, is a method for the synthesis of petroleum products. The starting material may be oxygen from the air and almost any hydrocarbon material, such as natural gas, coal, heavy crude oil or fuel oil. Due to its ability to use sources of carbon other than from petroleum to manufacture products such as motor fuel, the Hydrocol Process may be looked on as a means of greatly extending crude oil resources. Development of the process using natural gas as the raw material has brought it in line competitively with conventional refining processes, such as catalytic cracking, for the production of high-grade motor fuels. A number of large commercial installations are now under way.

For a number of years, particularly during the past war period, the Germans prepared a synthetic gasoline from carbon monoxide and hydrogen by the Fischer-Tropsch process. In this operation the Germans reacted steam with coke and then adjusted the ratio of hydrogen to CO to that desired in the synthesis step by passing the gas through a water-gas shift reactor which converted some of the carbon monoxide to carbon dioxide and hydrogen. The resulting gas freed from its CO_2 was then passed over a cobalt catalyst in a fixed bed. This process had many shortcomings:

1. The catalyst was expensive and reworking was frequent.
2. The catalyst was extremely heat sensitive; this coupled with the fact that enormous quantities of heat are liberated in the reaction made for complicated and expensive reactor design.
3. Power recovery from the reaction was low.
4. The gasoline produced was of poor quality (octane number 40 ASTM).

These drawbacks were so serious that the Fischer-Tropsch process could be justified economically only under an artificial economy. It appeared, however, that the basic reaction of CO and hydrogen to produce gasoline and other hydrocarbon products had a definite place in American economy. With that in mind, chemists and engineers of Hydrocarbon Research, Inc., and associates have been studying the process for more than eight years. The Hydrocol Process, as offered for license today, represents the culmination of these years of work. Its achievement depended upon the successful solution of the following problems:

1. The manufacture of cheap oxygen.
2. The efficient production of Hydrocol or synthesis gas from natural gas and oxygen.
3. The development of a cheap, rugged and efficient catalyst to promote the conversion of Hydrocol gas into gasoline.
4. The engineering of an integrated plant, including a reactor design, which would permit absolute temperature control and an efficient recovery of the heat of reaction in the form of useful power.

A simplified flow diagram of the process as applied to one of the plants now under construction is shown on the opposite page. Natural gas and oxygen separately preheated are caused to react in an open-type generator at 2500° F. and 250 psig. The resulting gas is scrubbed with water and then heated by exchange with the generator gas and divided into a major and minor stream. The minor stream is passed over a water-gas shift catalyst to convert the CO to hydrogen and CO_2. The CO_2 is scrubbed out and the hydrogen is passed to the catalyst treating plant. The major stream is introduced into the synthesis reactor where it is caused to react in the presence of a stream of recycle gas at 650° F. and 245 psig over a fluidized bed of catalyst. The products leaving the reactor are passed to a cyclone separator from which catalyst fines are recovered. These products pass through a heat exchanger where they are employed to heat recycle gas. From the heat exchanger the products pass to a cooler and into a separator. In the separator, separation into three phases occurs, namely gas, hydrocarbon and water. A portion of this gas is passed to an absorber, while the remainder is recycled back to the reactor. The water containing by-product oxygenated compounds is drawn off the bottom of the separator and treated for the removal of the chemicals, consisting chiefly of alcohols, ketones, acids and aldehydes. The small percentage of by-product oxygenated compounds contained in the hydrocarbon fraction is removed and the product is then fractionated. The off gas is passed to an absorber with gas from the separator, and the total C_3s and C_4s are polymerized and the polymer gasoline is combined with the treated gasoline. The unabsorbed gas is used as fuel for the oxygen and natural-gas preheaters. In the synthesis step 98.8 percent of the CO and 95.1 percent of the hydrogen are converted with yields as shown on the diagram.

Representative tests on Hydrocol gasoline blend are shown below.

Gravity, ° API	66.4
CFRM Octane, Clear	79.2
CFRR Octane, Clear	90.9
RVP at 100° F.	9.2

This brief description and accompanying flow diagram of the Hydrocol Process was made possible through the courtesy of Texaco Development Corporation, 135 East 42nd Street, New York 17, New York, licensors of the Hydrocol Process.

THE HYDROCOL PROCESS
SYNTHESIS GAS GENERATOR
STEAM
WATER
NATURAL GAS
75,000,000 STD. CU.FT./DAY
OXYGEN
45,000,000 STD. CU.FT./DAY
OXYGEN PLANT
AIR
CO CONVERTER
CO_2 ABSORBER
HYDROGEN TO CATALYST TREATING PLANT
SCRUBBER
CHARGE GAS
MAKE-UP WATER
REACTOR
STEAM
WATER
CYCLONE SEPARATOR
SEPARATOR
RECYCLE GAS
CHEMICAL RECOVERY
TREATING
FRACTIONATOR
GAS RECOVERY
FUEL GAS
POLYMERIZATION
PRODUCTS
GASOLINE BARRELS/DAY 6028
HEAVY OIL BARRELS/DAY 1138
OXYGENATED
CRDS-LB/DAY 200,000
COURTESY, TEXACO DEVELOPMENT CORPORATION

Butadiene Process

STANDARD OIL DEVELOPMENT COMPANY

RAPID development of commercial processes for the manufacture of butadiene resulted from the requirements of the wartime synthetic rubber program. Butadiene by the catalytic dehydrogenation of normal butane or butylene previously had been the subject of considerable research and pilot-plant work. In going from butane to butadiene it appeared that there would be some advantage in operating the process in two stages, while from butylenes a single-process stage was possible. Wartime demands of butylenes for gasoline production by alkylation and polymerization processes were heavy, but large-scale commercial development of catalytic cracking promised sufficient yields of unsaturates to meet requirements, and finally determined the choice of the one-stage route, using butylenes as the feed stock.

In the catalytic dehydrogenation of butylene, the principal process problem centers about the reactor. The design is complicated from the engineering standpoint by the requirements imposed by the chemical nature of the reactant and product, and by the thermodynamics of the reaction. The most serious limitation in process design is the tendency for butadiene to polymerize and ultimately to degrade to coke and hydrogen. This can be minimized by operating at hydrocarbon partial pressures of the order of 90 to 100 millimeters. The low partial pressures can be attained by low-pressure operation or by the use of an inert gas as a diluent.

To obtain reasonable butylene conversion, reaction temperatures above 1100° F. have been found necessary, but to avoid thermal cracking of the butylenes in heaters and transfer lines temperatures should not be in excess of 1300° F., so that for practical purposes the operating temperature range is 1150° to 1250° F. at a superficial reactant residence time of about 0.2 seconds. Because of the above considerations, it is necessary to carry out the reaction at about 25-30 percent conversion per pass and at reduced butylene partial pressure in order to prevent butadiene degradation and excessive by-product formation. An additional problem is presented by the fact that the reaction is highly endothermic, requiring an external heat source of about 725 Btu/per pound of butenes converted.

One commercial solution of these problems employs steam as a heat conveying medium to provide sufficient heat capacity to reduce temperature drop through the reactor to a workable range, 50-75° F., and simultaneously lower butylene partial pressure to the desired range of 100 millimeters. Excessive butylene temperatures are avoided by superheating the steam diluent above the reaction temperature and mixing it with the hydrocarbon preheated to 100-200° F. below reaction temperature just before it contacts the catalyst. An important advantage is that steam reacts with coke to form water gas at the temperatures employed for dehydrogenating butylene, allowing regeneration of the catalyst by steam alone.

A catalyst stable and active in the presence of steam was a requisite for the steam-dilution process, dehydrogenation catalysts used in early development work not being satisfactory from this standpoint. Extensive laboratory and pilot-plant tests resulted in development of a suitable catalyst for this process, which also gave higher butadiene yields and lower by-product formation than previously known catalysts under similar conditions.

The dehydrogenation process as finally developed is indicated on the flow sheet, and utilizes two single fixed-bed reactors, with equal cycle times on reaction and regeneration. Butylene from storage is preheated by means of heat exchange and a furnace to not over 1100° F., while the diluent steam is superheated to some 1300° F., and jet-type mixers are employed for contacting the two streams directly over the reactor-catalyst bed. A 25-pound pressure drop is taken across the mixing nozzles to insure adequate mixing energy.

Reaction products are quenched to below 1000° F. as they leave the catalyst bed by means of a pumped stream of condensate. The hot reactor effluent is further cooled in a waste-heat boiler and a series of quench towers, a condensate containing the bulk of the C_4 hydrocarbons finally being charged to a fractionator for stabilization by separation of a light overhead stream and to a rerun tower where a polymer bottoms stream is produced and the crude butadiene is condensed as an overhead product in conjunction with other C_4 hydrocarbons, this mixture constituting the feed to a subsequent purification process step where commercial butadiene is produced by extraction with cuprous salt solutions, or by other means.

Non-condensable vapors from the cooled reactor effluent condensate separator are passed through an absorber for recovery of C_4 hydrocarbons. Oil from the reactor effluent quench system is in a closed circulation system, but is maintained in satisfactory condition by means of a small make-up feed stream and a lean-oil stripper which continuously eliminates a small bottoms stream to slop. Reactor effluent polymer from the rerun tower is also rejected to slop. Steam generated in the waste heat boiler is utilized as dilution steam feed to the steam superheater furnaces, in conjunction with high-pressure steam from outside sources. Water condensed from the reactor effluent in the water separator is stripped of any volatile hydrocarbon contaminants in a water-stripping tower and is then discharged to a water-blowdown system.

The above data and the accompanying flow diagram are available here through cooperation of Standard Oil Development Company, New York.

BUTADIENE PROCESS
COURTESY, STANDARD OIL DEVELOPMENT COMPANY
BARRELS PER DAY
GALLONS PER MINUTE
POUNDS PER HOUR
MOLS PER HOUR
STANDARD CUBIC FEET PER DAY
PRESSURE, POUNDS PER SQUARE INCH
TEMPERATURE, DEGREES FAHRENHEIT
WATER COOLER
CHILLED WATER COOLER
REACTOR
OIL QUENCH TOWER
1000-1100
1250-1350
1150-1250
4-STEAM FURNACES
1-BUTENE FURNACE
HIGH PRESSURE STEAM
CONDENSATE QUENCH
WATER QUENCH TOWER
WATER SEPARATOR
WASTE HEAT BOILER
WATER
SUCTION K.O. DRUM
WATER STRIPPER
INTERSTAGE K.O. DRUM
LEAN OIL STRIPPER
STEAM
RECYCLE BUTADIENE FROM RUBBER PLANT
ABSORBER
STABILIZER
K.O. DRUM
GAS TO FUEL
RERUN TOWER
STEAM
STABILIZED C4 CUT
UNSTABILIZED C4 CUT
POLYMER
CONDENSATE SEPARATOR
FAT OIL
PURGE OIL TANK
SLOP
WATER TO BLOWDOWN SYSTEM
SPHERE
EXTRACTION SECTION FEED
MAKE UP LEAN OIL
SURGE
2-SPHERES
DEHYDROGENATION SECTION FEED

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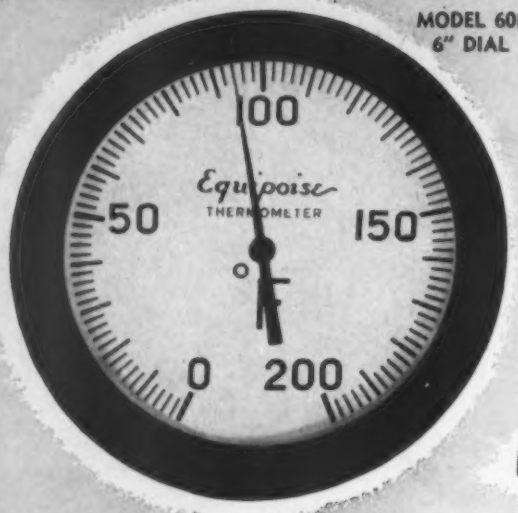
Primary Element—Precision bimetal helix.
Stem—18-8 stainless steel, ¼" O.D.
Special—One-piece connection nut and stem—machined from SOLID BAR STOCK, available in 2½" and 3½" stem lengths only. Stem ¼" O.D. Exclusive Equipoise feature!
Connection Nut—18-8 stainless steel, ½" SPT.
Over-Range—Ranges 500° F. and below—50%. Ranges over 500° F. —10%.
Accuracy—Within 1% of scale range.

MODEL 60F
6" DIAL



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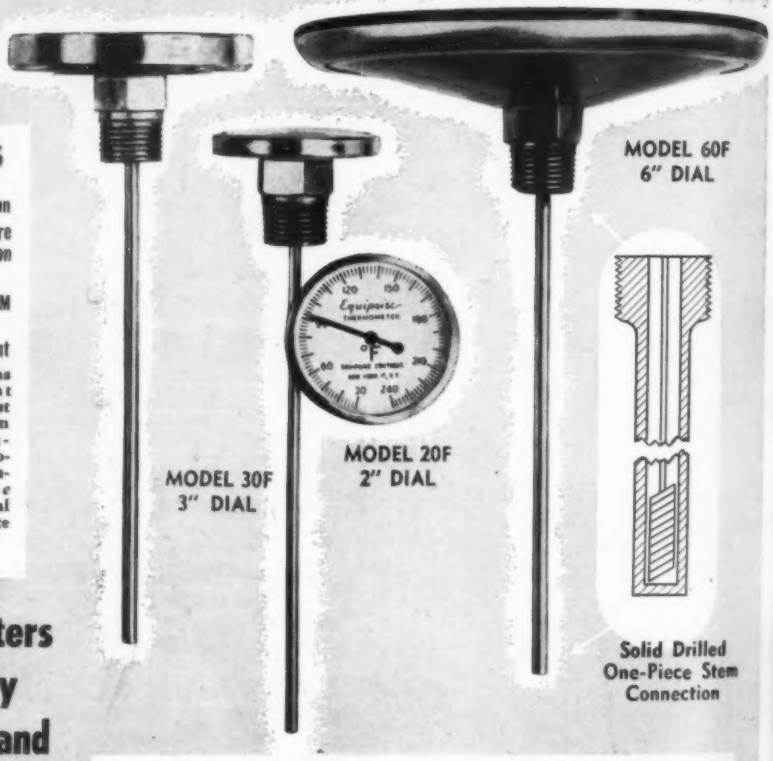

MODEL 60F
6" DIAL

MODEL 30F
3" DIAL

MODEL 20F
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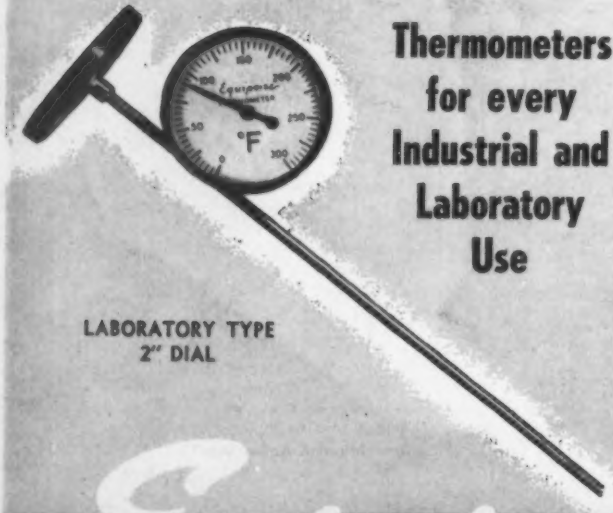
Solid Drilled
One-Piece Stem
Connection



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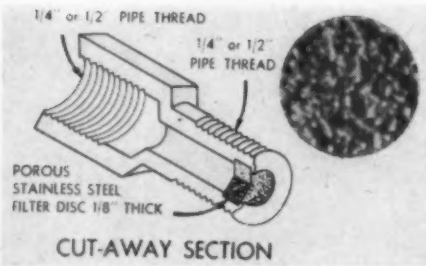

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